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Cover Image: The Suhar Refinery Improvement Project, Oman. Picture courtesy of Petrofac.

99TH ANNIVERSARY
|1922–2021|

HP Industry Perspectives

Asia continues to dominate new project market share

According to Gulf Energy Information's Global Energy Infrastructure database, new project announcements have decreased by approximately 6% year-over-year. However, new project announcements and delayed project restarts have gradually started to increase this year, as COVID-19 lockdown restrictions have been eased and more individuals get vaccinated.

Over the past year, the Asia-Pacific region represented 46% of new project market share (**FIG. 1**), followed by the U.S. and Eastern Europe, Russia and the Commonwealth of Independent States (CIS).

In Asia, China and India dominate new project announcements over the past year. In total, China (40%) and India (26%) represent 66% of new project announcements in the region. Both countries are investing heavily in new processing capacity to satisfy increasing demand for refined and petrochemical products, as well as in new LNG import and natural gas distribution infrastructure.

A major trend within the Asia-Pacific region is the investment in new refining and petrochemicals integrated complexes. China and India are investing heavily in these integrated complexes to satisfy increasing petrochemicals demand. For example, China is investing tens of billions of dollars in integrated facilities to significantly boost petrochemicals production. Several other Asian nations are doing the same (e.g., India, Indonesia, Brunei, among others).

In Europe, most capital projects are in Russia. The country continues to invest heavily in its refinery modernization program, expanding polymers production and increasing LNG export capacity.

The Middle East has announced several new projects over the past year to expand regional production capacity for clean fuels and petrochemical value chains.

The U.S. is continuing to build out its domestic petrochemicals industry and LNG export infrastructure, as well as boost its renewables fuels production capacity. HP

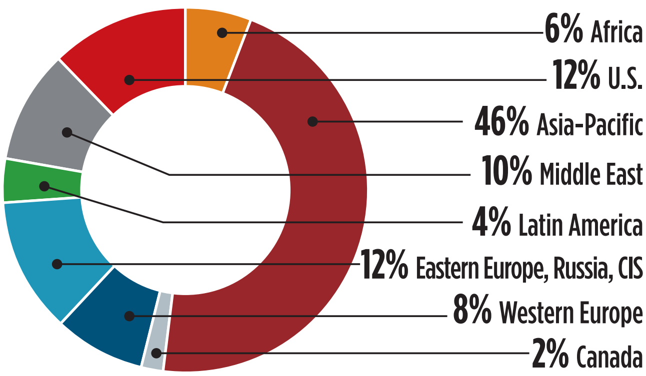

FIG. 1. New project market share by region, 2020–2021.

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HPI spending forecast to reach nearly $490 B in 2022

Each year, the editors of *Hydrocarbon Processing* provide the latest trends and spending forecast for the hydrocarbon processing industry (HPI). The *HPI Market Data 2022* provides HPI professionals with information, data, regulations/government initiatives, capital spending, supply/demand and advancing technologies that will affect the global processing industries.

Global lockdowns and travel restrictions due to the COVID-19 pandemic have taken a significant toll on refined fuels, petrochemicals and natural gas demand globally. However, as restrictions ease, demand and capital investments in new processing capacity are increasing. Where is demand/supply forecast to go? What kind of capital investments are being made in new processing capacity? What regulations and government/company initiatives will drastically affect future spending? These questions, along with major trends in the refining, petrochemicals, gas processing/LNG, maintenance and equipment sectors, are detailed and analyzed in *HPI Market Data 2022*.

Spending. In 2022, the editors of *Hydrocarbon Processing* forecast capital, maintenance and operating spending to reach $486 B (**TABLE 1**). Many capital projects that were delayed during COVID-19 restrictions have come back online. Gulf Energy Information's Global Energy Infrastructure (GEI) database, which tracks capital project construction in the hydrocarbon processing and renewables industries, has witnessed an uptick in positive final investment decisions, as well as the restart of several capital-intensive projects worldwide. However, due to the volatility of the global HPI, many projects are still being delayed, while a few have been abandoned.

In 2022, global maintenance and operating expenditures—excluding feedstock costs—are forecast to increase. Although more than 1 MMbpd of refining capacity was taken offline this year, new grassroots facilities and large-scale integrated complexes have begun operations. In turn, operating expenditures are likely to increase with this new capacity coming online. In addition, utilization rates have begun to increase in several regions, increasing operating expenditures.

Active project market share. The GEI database is tracking nearly 1,100 active HPI projects around the world. In total, these investments represent more than $1.8 T in capital expenditures to 2035. At 42%, the refining industry holds the largest active projects market share, followed by the petrochemicals sector (34%) and the gas processing/LNG sector (24%). The Asia-Pacific region represents the largest market share in both active projects and capital expenditures.

HPI Market Data 2022. To learn more about the major trends and initiatives that are affecting spending in the global HPI, the editors invite our readers to view an on-demand version of *Hydrocarbon Processing's* forecast webcast. A detailed analysis on capital projects, regulations/initiatives and supply/demand by country is available within the *HPI Market Data 2022* report, which can be purchased at www.store.gulfenergy.com. HP

TABLE 1. 2022 worldwide spending by budget, $B

	U.S.	Outside U.S.	Total
Capital	30	212	242
Maintenance	8	60	68
Operating	44	132	176
Total	**82**	**404**	**486**

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Executives' focus on lean management leaves companies exposed to risk during crises

In the wake of an extraordinary year marred by the impact of the COVID-19 pandemic, the author's company released the results of its 2021 Operational Risk Management Survey[1] of executives, a barometer of the topics that dominate boardroom discussions around the world.

The survey found that corporate executives' long-standing preoccupation with cost optimization within their organizations consequently had a negative impact on their companies' business resilience and agility during the COVID pandemic, as overly lean systems choked off their operational and environmental, social and governance (ESG) risk response options. This is also putting them at higher operational risk during another a crisis.

The pandemic presents executives with an ideal opportunity to change their approach toward operational excellence, which requires achieving a better balance between lean management, ESG and operational risk management to ensure agility, sustainability and resilience of operations in the future.

Cost reductions leave companies exposed to risk. Lean management has been the ultimate goal of large-scale organizations for the last decade or more. Around the world, companies have focused on minimizing their inventories, streamlining their supply chains and obtaining the highest possible productivity (and profit) from available resources and assets. Companies have realized higher margins by sourcing from low-cost labor markets, by implementing just-in-time manufacturing, and from other rationalizations and efficiencies.

During the pandemic, improving productivity and cost-efficiency continued to be the prime concern of the more than 200 senior executives polled for the survey. However, that focus had unforeseen consequences. With little room to maneuver, organizations did not have the flexibility to absorb the supply, sourcing, operating and commercial shocks caused by the pandemic.

More than half the organizations surveyed had to shut down operations during the pandemic, and a quarter struggled to get back on their feet or did not start up again at all (FIG. 1). Even though four out of five executives believed they had adequate crisis response planning in place, they still felt insufficiently prepared for the effects of a pandemic.

Ineffective anticipation of business threats and lack of recovery planning left many companies struggling to make a comeback. Although 81% of executives said they had a plan in place that they believed could address an unexpected business disruption, only 41% said that one year after the start of COVID they had been ready to respond to the pandemic. This means that effective crisis planning had clearly been insufficient among half of executives participating in the survey.

Asked how the risk profile of their organization had changed during the

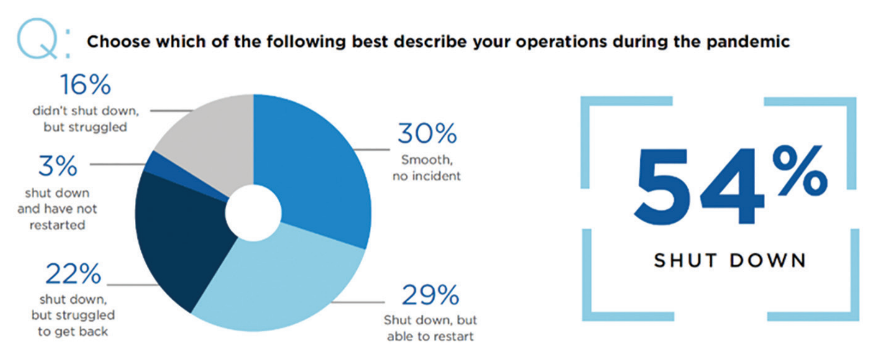

FIG. 1. Short-term thinking about risk is a significant threat.

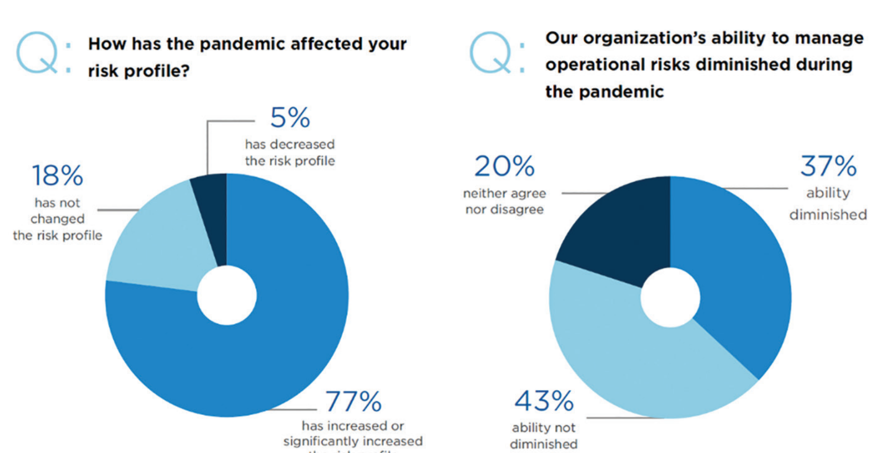

FIG. 2. Change in company risk profile during the pandemic.

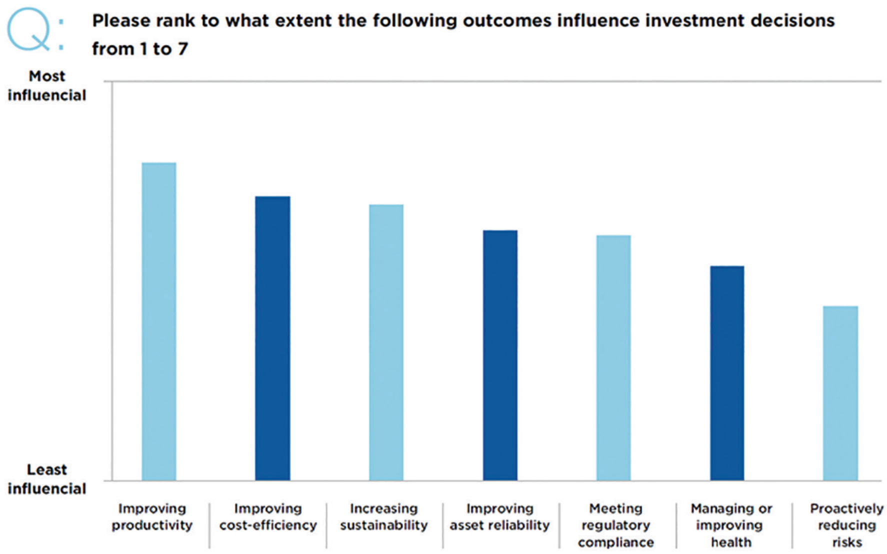

FIG. 3. Outcomes that drive investment decisions.

FIG. 4. An integrated approach may be the new paradigm for successful organizations.

pandemic, three out of four respondents predictably said it had increased (FIG. 2). More than half of the organizations that said their risk profile had increased were companies where productivity was the dominant topic at the boardroom level. Furthermore, of those organizations where health and safety ranked lowest on the boardroom agenda, 58% noticed a rise in their risk profile.

While executives pursued responses to various impacts caused by the pandemic, such as supply chain disruptions, lack of raw materials and workforce shortages, these responses tended to have a short-term focus and decisions taken in haste were based on fuzzy data. A long-term risk reduction strategy and recovery plan appear to have been missing.

Sustainability has gained boardroom attention. Considering all the challenges faced by senior executives during the pandemic, one stand-out topic that remained high on the boardroom agenda was sustainability.

Although operational excellence issues still get the lion's share of executives' attention in terms of cost-efficiency and productivity, head count and asset reliability discussions, sustainability ranks above issues such as regulatory compliance and proactive risk reduction (FIG. 3). ESG—and by extension sustainability—has earned its place on the management agenda, likely due in part to regulations and stakeholder and shareholder pressure, but is also symptomatic of the role ESG plays in determining the future ability of organizations to survive and operate.

Effective ESG management requires consideration of environmental and social issues and risks that hyper-lean organizations struggle to consider and include. Without a more balanced approach that examines cost optimization, ESG and risk management holistically, environmental or social disruption is likely—sooner or later—to affect business resilience.

Improve operations excellence through better risk management. Organizations have survived the pandemic partly due to government aid, and partly because their competitors were in the same boat. Had only a single organization or region been struck by the pandemic while competitors remained untouched, businesses would have found it much harder to bounce back and maintain their market position. The next crisis may not impact everyone the same way and businesses probably will not get the same level of support they received this time around.

In crises, most organizations operate at or close to their maximum risk limit. Proactively looking at short-, medium- and long-term future vulnerabilities is a way to anticipate and mitigate the consequences of emergencies. Many risks may not surface for years because of something organizations do today. Accurate risk profiling and data-based processes can help to identify current and future critical threats and determine an organization's level of risk tolerance. Instead of just reacting to unforeseen events, the development of a contingency plan and targeted controls built on data-driven risk profiling will enable organizations to make informed decisions during crises, so they can transform risks into opportunities without compromising long-term sustainability.

However, this requires a reevaluation of lean management and operational excellence. There are some signs this is already underway. As the 2021 survey shows, sustainability is already a topic that is high on the management agenda. This may indicate a shift towards a more sustainable model of operational excellence that factors in ESG considerations. Future investments and stakeholders will, after all, judge organizations on their alignment with ESG goals, and management of ESG risks will also determine the resilience of organizations.

Environmental issues have taken some time to make it into the boardroom, but they are there now. Perhaps the lessons learned from the operational risks posed by the pandemic may spell the end of traditional lean management practices and lead to a more balanced, less siloed view of operations. It is obvious that lean management alone is not sufficient to safeguard companies in all situations. The new paradigm for successful organizations looks more likely to be an integrated management approach that balances operational excellence with ESG and operational risks to ensure agility, sustainability and resilience of operations (FIG. 4). HP

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Disregarding storage preservation has consequences

As a rule, equipment to be installed at a plant site initially arrives at an outdoor storage yard. **FIG. 1** shows a small section where, on occasion, more than 1,000 items of rotating machinery are stored under slight oil mist pressure.

FIG. 1. Outdoor storage yard. Source: T. F. Hudgins.

Experience shows that of 1,000 machines stored for 3 yr in this manner, no more than 60 have ever shown incipient bearing damage. With ambient air and atmospheric contaminants excluded, it can be reasoned that the bearings had slight damage at assembly or sustained damage while being transported over roadways with potholes or on railroads with discontinuities in the track system. Such discontinuities can transmit shock loads into the cargo in the railcars.

Be this as it may, **FIG. 2** shows that fully 94% (or an even higher percentage) of the equipment properly preserved by oil mist will be ready to run up to 3 yr later without first having to be dismantled for cleanup.

FIG. 2 also allows us to estimate that after 3 yr of conventional preservation, the owners can expect a 20% "infant mortality." Equipment arriving at an outside storage yard for oil mist preservation receives an oil mist inlet fitting and a drilled drain plug. Once a small oil mist line is connected to the inlet fitting, the equipment can be stored for years of successful outdoor preservation.

Summary of how oil mist works. Oil mist will coalesce when atomized droplets combine to become larger drops of liquid oil as the equipment operates. Shaft rotation at operating speed causes turbulent flow in the bearings; the atomized particles get knocked together and become large and heavy. Turbulence is obviously not created in equipment at standstill. Since there is then no turbulence, the reclassifying (conversion) of oil mist to liquid oil proceeds at a slower rate in nonrunning equipment. However, since the oil mist is at a pressure of between 0.1 psi and 0.3 psi over ambient, it immediately fills the bearing housings and other spaces. Full corrosion protection is, therefore, established without delay. After a week or so, all surfaces will show a light coating of oil.

Noteworthy facts include:

- A low-point drain location is identified and left open at all times. This low point is useful because it prevents oil accumulation and ensures throughflow of a small amount of mist. The mist lost through a drain orifice equals the makeup mist. With rare exceptions, a 1-mm minimum, 3-mm maximum drain orifice is all that is needed.
- A machine in storage and ready for preservation can be slanted or skewed to create or locate a convenient and/or accessible low-point drain. A flat pan can be used to catch this drainage for metering and proper disposal.
- A rotor blanketed with oil mist should be manually rotated 2.5 turns every 6 mos. This ensures that a light coating of oil remains established throughout a rolling element bearing and avoids false brinelling (indentation) at or near the 6 o'clock load points.

Only steam turbines require special cleanup before recommissioning. Accordingly, the following precautions apply only to steam turbines:

- Use oils with formulations that will not promote future stress corrosion cracking.
- Prior to machine commissioning, blow steam through the equipment. This removes the thin coating of oil that would have accumulated on the surfaces.
- Collect some of the exiting blowdown steam and observe if it still contains traces of oil.

Fifty years of oil mist preservation. Although oil mist preservation has been used for every successful project at best-in-class companies and corporations, tra-

dition and seriously flawed decisions can impede progress. During an economic downturn circa 2008, a corporate headquarters organization demanded immediate cancellation of a major project. Immediate cessation of all work was decreed, although cancellation required abandoning work on an almost-completed oil mist storage yard full of rotating equipment. The fluid machines had only recently arrived in anticipation of being installed on close to 100 foundations.

The signal to resume work came years later, and the ultimate cost of the "leave unprotected mandate" flew in the face of every aspect of reasonable risk management. Millions of dollars were spent in unavoidable dismantling, parts replacement and equipment reassembly. Today, a 9-gal oil mist console lists for less than $50,000, and chances are it cost considerably less in 2008. This amount would have included a mist density monitoring instrument and a back-mounted spare. To be fair, let us add $50,000 in today's dollars for labor and temporary, 1.5-in. field piping—still pocket change compared to rebuilding a massive amount of equipment that had been left unprotected in weather conditions that ranged from –20°F to +105°F and included snow, ice, torrential rains and blistering heat.

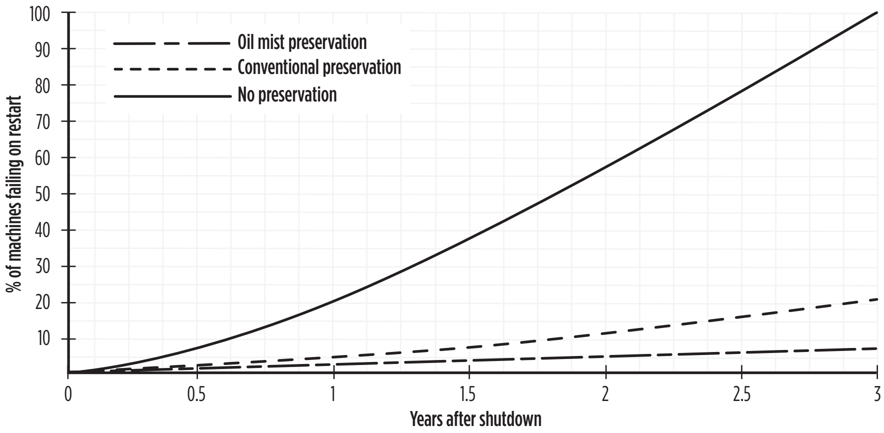

FIG. 2. Percentage of defective machines vs. preservation method and elapsed time.

We are now in late 2021, and a strangely related experience is making the rounds of a worldwide network of reliability professionals. It involves a major project where an EPC firm will not be able to use oil mist in an outdoor staging area, which is where some of the equipment will be stored for years. Apparently, the EPC's contract with the equipment owner requires that every one of probably hundreds of machines will be dismantled and cleaned before being reassembled and ultimately commissioned.

If these unconfirmed rumors were true, three undisputed facts would apply:

1. Storage with oil mist blanketing would have saved the owner an inordinate amount of money—perhaps 100 times the cost of dismantling the machines, followed by replacing corroded parts and reassembling subassemblies and full assemblies.
2. The storage with oil mist blanketing would totally inhibit corrosion, and the above tasks could probably be done at one-tenths of the likely cost.
3. This again shows that some management members never have time to obtain facts on what reliability improvement really entails, and what potential cost savings are lost to both the company and its stakeholders. HP

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Improved methods for trace element analysis of challenging petrochemical samples

There is no doubt that the purpose of refining crude oils into natural raw materials is crucial to produce a vast array of fuels and products used in our daily lives. However, impurities can often enter the refinement process and reduce the quality grades of crude oil extracts, intermediates and final products. Typical impurities, such as metals (e.g., vanadium, mercury or lead), are known as catalyst poisons. These can severely impact the performance and value of the final product and have detrimental effects on the petroleum refining processes. These metallic impurities can affect products even at very low levels and have the potential to cause interruptions in the production cycle by causing equipment corrosion, catalyst poisoning and build-up in deposits in engines.

Other elements like arsenic or sulfur are released as toxic emissions when fuels are combusted. Sulfur is a common impurity in crude oil. Sour crude oils with high quantities of sulfur compounds must have sulfur extracted to improve the quality of the oil, as well as to adhere to air quality requirements.

To help reduce the presence of impurities in petrochemical products, more stringent industry and environmental standards mean oil samples must be analyzed with high accuracy from crude, intermediate and final product stages of refinement.

Considering these challenges that impurities cause to the petrochemical industry, it is critical for laboratories to monitor the levels of impurities (e.g., metals and sulfur) in common petrochemical products with highly sensitive methods. This article will discuss how advances in elemental mass spectrometry systems are enabling rapid and accurate element analysis of impurities, and how small changes in the sample introduction system—such as the use of ceramic torches—are further boosting operational capacities.

The challenges of trace element analysis. In many laboratories, the analysis of crude, intermediates and final products for trace elements is accomplished by using inductively coupled plasma optical emissions spectrometry (ICP-OES) or mass spectrometry (ICP-MS). ICP-OES is a highly flexible technique that can perform fast and robust analyses of samples. ICP-MS is less commonly used, although it offers detection limits improved by several orders of magnitude and enables identification and subsequent quantification of sulfur containing compounds when it is coupled with a suitable separation technique, such as gas chromatography (GC).

Testing laboratories face several challenges when monitoring petrochemical impurities. First, the variability of physicochemical properties of petrochemical products means dedicated method workflows for each sample type are often required. Different viscosities of petroleum and many of its derivatives make it difficult to analyze multiple sample types directly by ICP-OES or ICP-MS using a single method.

For example, organic samples must be aspirated into plasma for analysis, and several steps must be taken to prevent carbon build-up and minimize background interferences. High carbon is problematic for analysis, as it may deposit on the system and can cause interferences on analytes, such as chromium and vanadium. Combustible organic samples (e.g., naphtha or aviation turbine fuel) have particularly high carbon content, an issue that increases the potential for carbon deposition in the torch and interface region of mass spectrometer systems.

A major challenge faced by using ICP-OES is the limit of detection (LOD) possible for impurities. ICP-OES usually has a LOD in the parts per billion (ppb) range, whereas ICP-MS can achieve parts per trillion (ppt) when detecting analytes. While this higher sensitivity of ICP-MS is beneficial, the method can have challenges in measuring key elements, such as phosphorus and sulfur at low levels because they are strongly interfered.

Improving testing quality and efficiency with triple quadrupole ICP-MS workflows. New advances in ICP-MS technologies are helping laboratories to overcome many of these challenges in petrochemical analysis. Advanced ICP-MS technologies are greatly improving detection limits and lowering interference levels for elements, like phosphorus and sulfur. ICP-MS systems also rapidly decrease analysis times, allowing higher throughput analysis of multiple samples. ICP-MS-based methodologies are also easy to learn, with straightforward method set-up and clear data visualization, meaning that laboratories can train scientists quicker for petrochemical analysis.

Single quadrupole ICP-MS provides collision cell technology with helium gas and kinetic energy discrimination (KED), which enables many argon- and carbon-based polyatomic interferences to be removed. While single quadrupole ICP-MS systems generally fulfil the requirements in terms of sensitivity and selectivity, they often have problems analyzing sulfur and phosphorus due to high levels of interference.

Triple quadrupole ICP-MS systems provide improved accuracy and detection limits for elements challenged by severe interferences. The solution lies with an alternate method of interference removal. With single quadrupole systems, interferences with sulfur are not fully removable using helium and KED. However, triple quadrupole ICP-MS uses oxygen (O_2) as the cell gas to convert sulfur to $^{32}S^{16}O^+$, thereby, moving it away from the very large $^{16}O_2^+$ interference and leading to dramatically improved detection limits.

While single quadrupole ICP-MS is typically capable of measuring sulfur concentrations in the high ppb or even ppt range, triple quadrupole ICP-MS systems can easily meet detection limits in the low or even sub ppb level. Moreover, triple quadrupole ICP-MS enables the use of other reactive gases like ammonia for interference removal, allowing the elimination of interferences affecting vanadium and chromium.

Greater robustness with the ceramic torch. While single and triple quadrupole ICP-MS systems are improving petrochemical analysis, there are also ways to boost the efficiency and cost-effectiveness of these systems by making small adjustments to their specifications. For example, implementing the use of the latest ceramic torches can improve the functionality of ICP-MS systems.

The torch is in direct contact with the plasma and sample aerosol generated by the nebulizer, making it a vital component for ensuring consistent and reliable ICP-MS operation. Conventional torches are made of quartz and require routine maintenance or frequent replacement—especially when performing analysis with crude oil samples—and, therefore, are significant contributors to running costs. However, ceramic torches have been used frequently with ICP-OES and show an improved product lifetime and reduced maintenance requirements, making them more cost-effective in the long-term vs. quartz torches.

Moreover, ceramic torches can enhance the quality and sensitivity of petrochemical analysis. While quartz torches can contribute to background signals, particularly for silicon, ceramic-based torches provide lower silicon background levels and deliver improved detection limits for silicon and other elements, such as cobalt and indium.

Takeaway. Regulatory requirements on monitoring increasingly lower levels of impurities in petrochemical products means that more advanced method platforms are required in laboratories. Single and triple quadrupole ICP-MS systems are boosting the capabilities of impurity monitoring in oil and refinery product testing, allowing sensitive analysis and comprehensive interference removal for all elements commonly analyzed in the industry.

Triple quadrupole ICP-MS systems are increasing analysis sensitivity levels by utilizing a variety of reactive gases—such as oxygen and ammonia—to expand analytical capability. This is especially true for elements, like sulfur, that are difficult to measure at lower levels with single quadrupole ICP-MS. New ceramic-based torches are further enhancing the quality of petrochemical analysis by providing greater robustness of mass spectrometry systems and improving detection limits of several elements. Overall, these systems offer new options to overcome challenges in petrochemical sample analysis and help to improve workflow efficiencies. HP

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Maintaining profitability with opportunity crudes while mitigating naphthenic acid corrosion

Refiners are asked to produce products within ever tighter margins in a changing industry. Operators that strive to increase their profitability often turn to cheaper but more corrosive opportunity crudes; however, that is not without risks.

These risks most often manifest as unplanned shutdowns caused by the detection of unacceptably high corrosion activity in a certain area of the plant. If unnoticed and unmitigated, this excessive corrosion could result in a hydrocarbon leak and in the worst case, an explosion or fire. Such an event may lead to human tragedy, extended business interruption, loss of custom, the cost of extensive rebuilding of equipment, negative impact to the company's brand reputation and future enhanced regulatory scrutiny.

A recent example of the danger of undetected corrosion was the 2019 explosion at a Philadelphia hydrocarbon facility caused by an elbow joint that had been installed in 1973 and had worn down to the point of failing. This disaster released dangerous chemicals in the air and ultimately bankrupted the company.

One of the biggest challenges—and corrosion culprits—when using cheap opportunity crudes is naphthenic acid corrosion, which is a high-temperature corrosion mechanism that commonly occurs in distillation columns. Since naphthenic acid corrosion starts occurring at temperatures greater than 250°C (482°F), it is often difficult to assess in real time and poses a unique challenge to the corrosion mitigation plans at a refinery.

What is naphthenic acid and why is it so corrosive? High total acid number crudes are known to bring high levels of naphthenic acid with them. Naphthenic acid corrosion is a particularly aggressive and often localized corrosion characterized by an 'orange peel' effect within pipes. Naphthenic acid is naturally occurring in certain types of crude brought up in oil exploration.

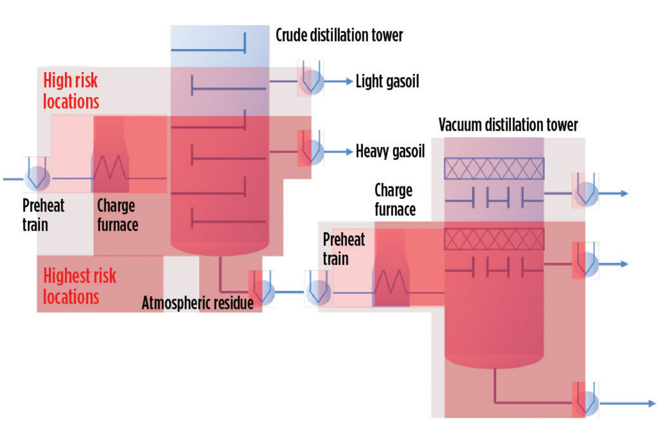

FIG. 1. Example of high-risk locations for naphthenic acid attack in the crude and vacuum distillation towers.

It only becomes a problem in areas of the crude and vacuum distillation units at temperatures above 200°C (392°F). Starting at the hot end of the preheat train, through the charge heater coils and affecting the lower section of the crude distillation tower, light and heavy gasoil side draw and residue product run down until they are cooled back below this critical temperature threshold (**FIG. 1**). The danger areas for naphthenic acid corrosion are the same in the vacuum unit.

Generally, naphthenic acid is most aggressive to carbon steel. Stainless steel has shown a greater resistance but would be prohibitively expensive to use, which is why most operators continue to use carbon steel.

The naphthenic acid molecules can be relatively inert until they are 'activated' by high shear velocities—such as those found at flow discontinuities caused by reducers and expanders, by elbows and bends and by tee sections, or downstream of any flow disturbances, such as a pump discharge, injection nozzles or intrusive probes. This makes these areas most susceptible to naphthenic acid corrosion.

Managing naphthenic acid corrosion. Remaining vigilant to the dangers of naphthenic acid corrosion in a facility demands having a plan in place to monitor on an ongoing basis. Traditionally, the most common types of monitoring used are intrusive corrosion probes or manual ultrasound. Both methods pose challenges. The corrosion probe provides an indirect measurement as the tips corrode off over time, thereby not providing an accurate picture. Manual ultrasound requires personnel to put themselves in potentially dangerous positions.

If the intrusive probe has a tip left, it will provide valuable and instant data about the pipe conditions. They can quickly alert an operator to a risk developing in a pipe. Still, it should be noted that the installation and replacement of such probes can introduce different types of safety risks.

Manual ultrasound presents additional problems in that the results are not reproducible. It is highly unlikely that the technician sent out to apply the manual ultrasound device to take a reading will take readings at the exact same spot on the pipe each time. This can lead to missed corrosion and an incomplete map of where corrosion might be developing. In addition, manual measurements are not practical to carry out frequently, so a quickly developing corrosion risk may go unnoticed.

In areas of the process that undergo extreme temperatures, corrosion monitoring becomes even more difficult, as instrumentation is not able to cope with the temperature and the extreme heat poses a safety risk to personnel.

A better solution. In recent years, the use of ultrasonic wireless measurement tools with a wider field of measurement have found an eager market. They offer reliable monitoring with low-cost devices that wirelessly transmit the data back to a central hub. The real benefit is that with low-cost devices, many more can be deployed across a facility, creating a network of corrosion data.

Extreme temperatures remain a challenge to obtain accurate readings from any device. However, a new solution was recently introduced to the marketplace that is non-intrusive, works electromagnetically, does not need a coolant and is suitable for use in all hazardous areas. This magnetic corrosion monitoring device makes it possible to create a network of data that more completely covers the facility. This network also boosts the digital transformation most facilities are implementing to help them keep up with demands.

Each sensor has a measurement footprint of approximately 1 cm^2, which is like the range of a manual ultrasound. Detecting corrosion from the reading of a single ultrasound sensor would be limited if there was only one installed. However, installing sensors as multipoint arrays at the highest risk locations significantly increases the probability of detecting corrosion. The high-risk locations are chosen, considering temperature, metallurgy, the equipment geometry and the acid boiling range distribution of the type of crude oil used.

Maintaining profit. Opportunity crudes are not going away. They are a way for a changing industry to remain profitable and productive. However, opportunity crudes pose challenges that could impact a plant and the bottom line. Meeting demands in the marketplace means becoming adaptable, while keeping a facility running safely and at optimal capacity. Establishing a wireless, non-intrusive corrosion monitoring network within a processing facility ensures a proactive approach to corrosion mitigation.

A quick, back of the envelope calculation shows the financial benefits of knowing the corrosion status in a plant. For example, take a medium-sized refinery and examine the cost of an unplanned piping replacement vs. a planned and well-prepared piping replacement. The cost of a planned replacement alone is estimated to save around $500,000. The estimated lost profit of an unplanned shutdown can quickly run up to around $300,000/d, depending on the necessary actions needed and assuming the repairs take a minimum of 4 d. Totaling the costs, it becomes clear that unplanned shutdowns can significantly impact profitability, sometimes ranging in the millions of dollars. It makes a compelling case for implementing a network of corrosion monitoring probes that map the facility in enough detail to avoid unplanned piping replacements. HP

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Choosing an FCCU CO promoter is no longer driven by NO_x emissions

Fluidized catalytic cracking units (FCCUs) form coke during the reaction that combusts in the regenerator and supplies heat for the process. The reactions can be full combustion to carbon dioxide (CO_2) and water or partial combustion to carbon monoxide (CO) and water emissions. CO can further combust to CO_2. An FCCU may utilize CO promoter additives to catalyze these reactions to:

- Meet environmental regulations on CO emissions
- Reduce afterburn, which occurs when CO and oxygen react in the dilute phase (or less commonly the flue gas line) of the regenerator. In the dilute phase, there is less catalyst to absorb the heat of combustion, potentially causing significant temperature increases above metallurgical limits.

Beyond the desired oxidation reactions, promoters also catalyze the unwanted reaction of nitrogen in the coke to form nitrogen oxides (NO_x) (vs. N_2). Depending on NO_x restrictions, there are two types of platinum group metals (PGM):

1. Platinum promoters that allow higher degrees of CO promotion but generate substantial NO_x
2. Non-platinum promoters, which typically use palladium that generate comparably less NO_x but are less effective in CO oxidation.

In early 2018, platinum and palladium both cost approximately $1,000/troy ounce (oz t). At present, palladium is more than double the price of platinum. As palladium-based catalysts tend to require more metal to achieve equivalent CO promotion, a palladium-based CO promoter is around $25/lb more expensive than a platinum-based CO promoter due to metals cost alone. For example, an average unit uses 2 lb–5 lb of CO promoter additive per sT catalyst. At 5 sT/d, that could be more than $200,000/yr higher costs to use a palladium promoter vs. a platinum promoter. **Note:** The metals are generally not considered recoverable. Given this, FCCUs will preferably use platinum-based CO promoters. Palladium is used by units that either have an environmental consent decree and/or cannot meet NO_x emissions with platinum-based promoters.

Novel promoters. Conventional promoters also have a short half-life (around a few days) due to metal sintering under high-temperature conditions in the FCCU. To mitigate these issues, a new generation of CO promoters has been developed. These novel promoters use a tuned PGM-support interaction, hindering sintering by anchoring PGM on the support. The resulting morphology modification (**FIG. 1**) improves promoter durability. Anchoring enables the ability to control PGM-oxygen interaction through oxygen on-demand shuttling. The products are designed to have optimized attrition, density and particle size distribution for use in the FCCU.

The co-authors' company has designed a novel, proprietary, durable platinum-based CO promoter[a] that generates significantly less NO_x vs. traditional platinum promoters. In laboratory testing, the proprietary CO promoter was compared to the latest generation of palladium and platinum promoters (**TABLE 1**). The proprietary promoter matched palladium promoter activity at 15% lower addition. The NO_x generation increases about 10% when using palladium—in comparison, the platinum promoter generated 160% more NO_x vs. the proprietary design. For most units, the proprietary CO promoter[a] allows the choice between palladium vs. platinum to no longer be driven by NO_x emissions constraints, but rather based on PGM pricing.

The first commercial adoption of the proprietary CO promoter was at CHS's Laurel refinery outside Billings, Mon-

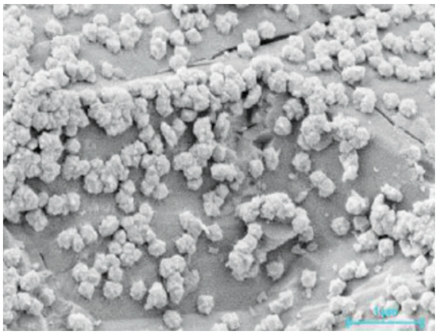

FIG. 1. The image, using a scanning electron microscope, shows the customized surface morphology of the CO promoter.

TABLE 1. Lab testing results for commercial palladium- and platinum-based promoters vs. the proprietary promoter[a]

	Palladium-based	Platinum-based	Proprietary promoter[a]
CO conversion, %	50	50	50
NO_x conversion, %	75	200	82

Note: Samples were aged and normalized on CO conversion (%) basis. Aging: 12h, T_{max} = 780°C; flow-through reactor: 10 min air, 10 min N_2, 10 min H_2, 10 min N_2, constant 10% H_2O-steam. Test: 99% spent catalyst + 1% promoter, 1 l/min, 2 vol% O_2/N_2, 700°C.

tana. The refinery processes a severely hydrotreated vacuum gasoil feed and operates in full burn. The site previously used a preblended palladium-based promoter[c] for NO_x control. The proprietary promoter replaced the palladium promoter at a preblended 12.5% lower addition rate to achieve similar CO oxidation activity, while staying within the desired NO_x limits. The afterburn decreased by half—from 6.67°C (44°F) to -5.56°C (22°F)—at a lower regenerator bed temperature (a shift to a lighter feed reduced bed temperature) (TABLE 2). Most importantly, the overall promoter cost ($/d) decreased—a desired benefit in the COVID-19 economic environment.

Takeaway. Modern platinum-based CO promoters have a price advantage over palladium-based promoters due to PGM pricing. However, NO_x limits can inhibit the adoption of platinum promoters. The next-generation, proprietary platinum-based CO promoter[a] generates less than 10% more NO_x vs. palladium promoters. The technology's custom-tuned PGM-support significantly reduces sintering for enhanced CO promotion durability. With this next-generation promoter, the choice between palladium or platinum promoters is no longer driven by NO_x emissions constraints but rather by pricing and performance.

In the first commercial trial at the CHS Laurel refinery, the proprietary platinum-based CO promoter replaced a palladium-based promoter. The new platinum-based CO promoter yielded similar CO emissions, easily stayed within NO_x constraints and reduced afterburn by 22°F, all at 12.5% lower addition rate. The proprietary platinum-based CO promoter enables refiners previously limited to high-priced palladium promoters to reduce operating expenses and improve performance. HP

TABLE 2. CHS Laurel regenerator performance for the palladium-based CO promoter[b] and the novel platinum-based promoter[a]

Promoter	CO, mol%	NOx, ppm vol	Catalyst bed temperature, °F	Dilute phase temperature, °F	Afterburn, °F
Palladium-based CO promoter[b]	0.005 ± 0.002	35.1 ± 7.3	1,310 ± 6.2	1,354 ± 8.1	43.9 ± 6.7
Platinum-based CO promoter[a]	0.006 ± 0.002	40.3 ± 8.2	1,300 ± 4.9	1,322 ± 5.2	21.9 ± 6.7

NOTES

[a] BASF's ENABLE™ CO promoter
[b] BASF'S CONQUERNOX®

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Establish the maximum allowable pressure drop across the catalyst bed structure in a hydrocracker

The reaction section of a hydrocracker unit has three reactors configured in two stages. Most of the cracking takes place in the third reactor (of interest), which uses a high-activity hydrocracking catalyst in its top four beds. As the feed and recycle gas mixture enters the top of the reactor, it flows evenly over the reactor cross-sections before flowing down to each of the catalyst beds. The catalyst is designed to remove organic metals from the feed and is graded to provide maximum surface area to collect pipe scale and other solid particles with minimal impact on reactor pressure drop, as shown in **FIG. 1.**

As per original equipment manufacturer (OEM) specifications, each catalyst bed can handle an 85-psig pressure drop. Pressure differential transmitters measure the differential pressure over the reactor and can also be lined up to indicate the pressure drop over the top catalyst beds.

A steady increase in pressure drop is typically expected across the first catalyst bed structure during normal operations (**FIG. 2**). In operational runs, it was observed that this pressure drop approached the set value of 85 psig much earlier than the scheduled end-of-run (EOR) conditions. Operational adjustments were required to avoid this pressure drop from exceeding an 85-psig set value before the scheduled EOR, which often compromised the efficiency and productivity of the hydrocracker.

Problem statement. To establish the maximum allowable pressure drop that the reactor internal structural components can withstand, the mechanical strength of the a) catalyst bed support structure, b) expansion skirt structure, c) shell integral support ring, and d) expansion skirt plug weld to the shell wall, must be established. A preliminary assessment of the reactor's mechanical design was completed, and several important characteristics of internal mechanical and fluid loads were identified for the existing maximum allowable 85 psid pressure drop permitted across the first catalyst bed and expansion skirt assembly, combined.

Basically, the dead weights for the reactor internals contributing toward the mechanical loads remain constant through the changing fluid loading scenarios. The correlation for pressure drop data across the various elevations of the reactor were obtained from the process trends of existing field instrumentation, which established that 86% of the pressure drop was found to occur across the catalyst beam structure of the first bed, while the remaining 14% was found to occur across the expansion skirt assembly of the first bed.

Among the contributions of the various fluid loads and the weights of the reactor internals contributing toward the mechanical loading of 398 psi on the top general surface of the shell integral support ring, the effect of the pressure drop across the first bed had the maximum impact of 94% when compared with the other. The contribution of the various fluid loads and the weights of the reactor internals toward the mechanical loading of 655 lb/in. (117 kg/cm) on the expansion skirt plug weld to shell is shown in **FIG. 3.** The effect of the pressure drop across the first bed had the maximum impact of 87% when compared with the other loads.

The complex geometry associated with the reactor internals, coupled with the thermo-mechanical loading, mandated an exhaustive and thorough stress analysis based on American Society of Mechanical Engineers ASME Sec-VIII Div-2 methodology.[1] The accurate stress distribution was simulated using the finite element method (FEM).

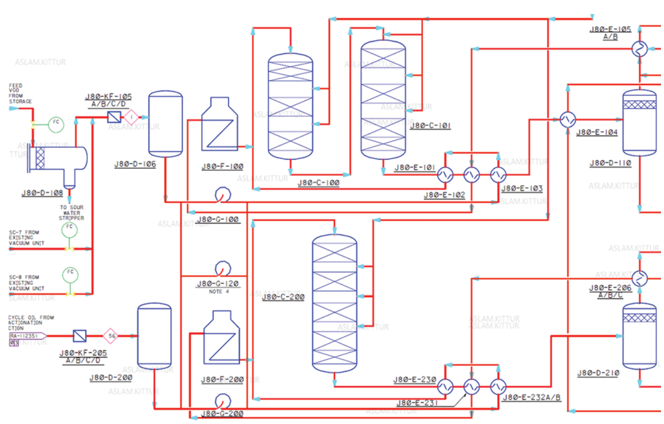

FIG. 1. Second-stage reactor in a hydrocracking unit.

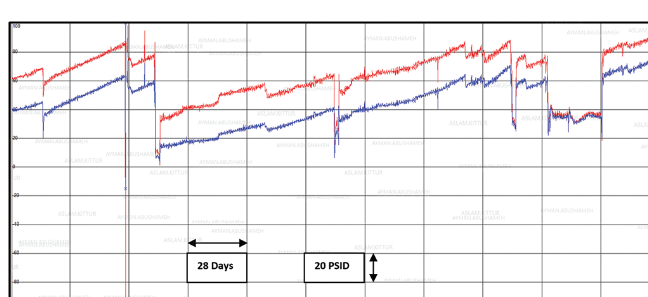

FIG. 2. Pressure drop process trends.

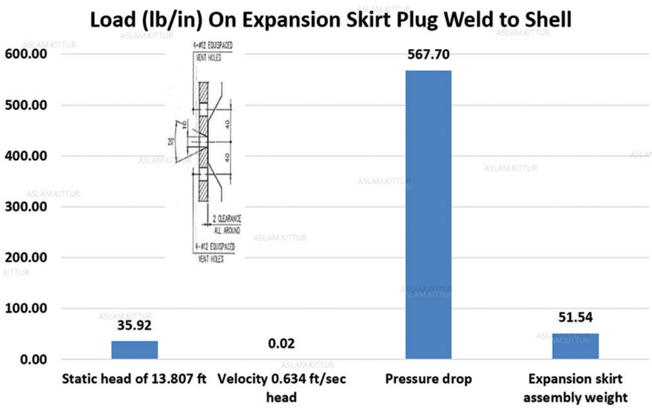

FIG.3. Load contribution on weld.

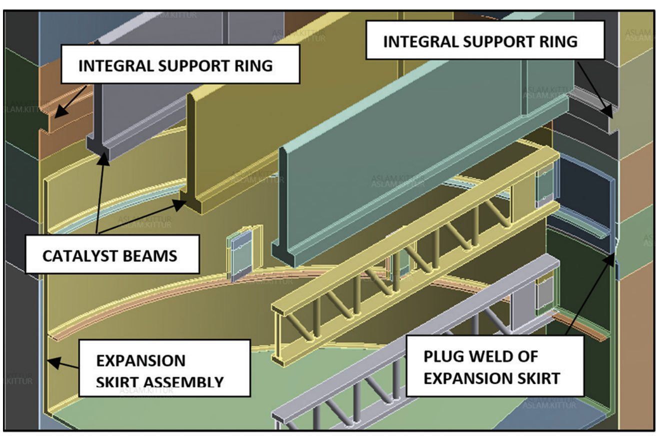

FIG. 4. Geometry of the FEM model.

Finite element modeling. The reactor was modelled as a 90° 3D-solid model based on the cyclical symmetry shown in **FIG. 4.** This model enables the accommodation of the geometric details for applying the appropriate loads and boundary conditions. The catalyst weight, self-weight of the support beams and the grids, operating liquid static head, velocity head, pressure drop mechanical loading and temperature distribution were considered while establishing the mechanical strength of the shell integral support ring. Whereas, for establishing the mechanical strength of the expansion skirt plug weld to shell, the self-weight of the expansion skirt and the lower liquid distribution tray, liquid collection tray assembly, mixing chamber, rough liquid distribution tray assembly, the operating liquid static head, velocity head and pressure drop mechanical loading on each of the three trays, and temperature distribution, were considered.

Similarly, the mechanical strength of the individual structural components of the reactor internals was also established based on individual weight, fluid and temperature distribution loads. The temperature-dependent mechanical properties for reactor shell base metal A336 GrF22, weld overlay A312 Gr347, and shell internals A312 Gr321 were obtained from ASME Sec-II Part-D.[2]

Analysis methodology. The reactor internal structural components were evaluated primarily for protection against plastic collapse. Despite the steady nature of operating pressure and temperature of the reactors, the ratcheting and fatigue failure modes were only considered due to the hydrocracker unit start-up and shutdowns cycles. The FEM with continuum elements provided the total stress distribution for evaluation and the simulated stresses were compared with stress limits defined in ASME Sec-VIII Div-2. This established the maximum load bearing capacity of the structure and, consequently, the maximum allowable pressure drop that the reactor can safely sustain.

Nomenclature used here includes:

P = Primary stress
P_m = General primary membrane equivalent stress
P_L = Local primary membrane equivalent stress
P_b = Primary bending equivalent stress
Q = Secondary equivalent stress
F = Peak stress produced by a stress conc. or a thermal stress over and above the nominal $(P + Q)$ stress level
S = von Mises equivalent stress
S_m = Allowable stress intensity
S_y = Yield strength
S_{alt} = Alternating stress intensity
ΔS_n = Stress intensity range.

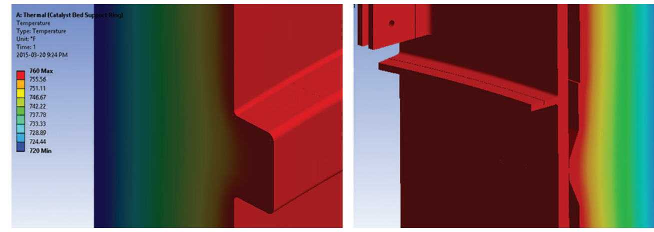

FIG. 5. Temperature distribution.

Temperature. A thermal stress was simulated for the reactor vessel as the thermal gradient along the radial direction was 40°F per 9 in. = 4.45°F/in., as shown in **FIG. 5.** The thermal gradient across the entire height (longitudinal) of the reactor was 8°F per 5 m = 0.04°F/in., which could be ignored as it was insignificant toward generating secondary stresses.

The FEM model in **FIGS. 6** and **7** shows the quality of the mesh on the stress classification planes (SCPs) and the location of the stress classification lines (SCLs) selected at the critical locations of gross and local structural discontinuities for this analysis. The FEM model has been meshed using hexahedrons (20-node brick elements), which can effectively simulate the stress and deformation fields in solids in a structural analysis. The stress components were obtained from the linear elastic stress analysis and were then integrated along the SCLs through the wall thickness to compute the equivalent linearized membrane, bending stresses and peak stress components.

Catalyst bed support ring at Bed #1 (**FIG. 6**):

- SCL-1 is the shell thickness at the general area (P_m and P_b)
- SCL-2, -3 and -4 are the gross discontinuity of the shell (P_L and Q).

Expansion skirt assembly at Bed #1 (**FIG. 7**):

- SCL-1 is the gross discontinuity of the shell (P_L and Q)
- SCL-2 and -3 are the concentration plug weld-root, connecting the top and bottom rings (F)
- SCL-4 is the local discontinuity of the plug weld connecting the top and bottom rings (P_L and Q).

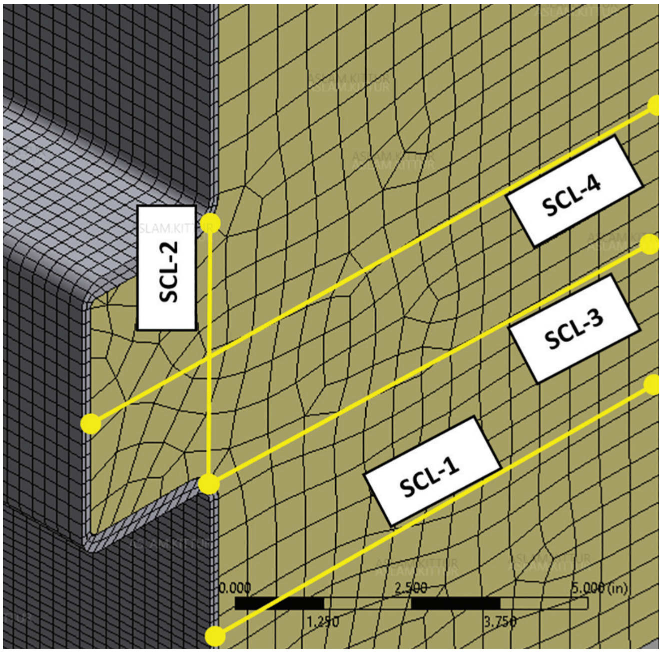

FIG. 6. SCL: Support ring.

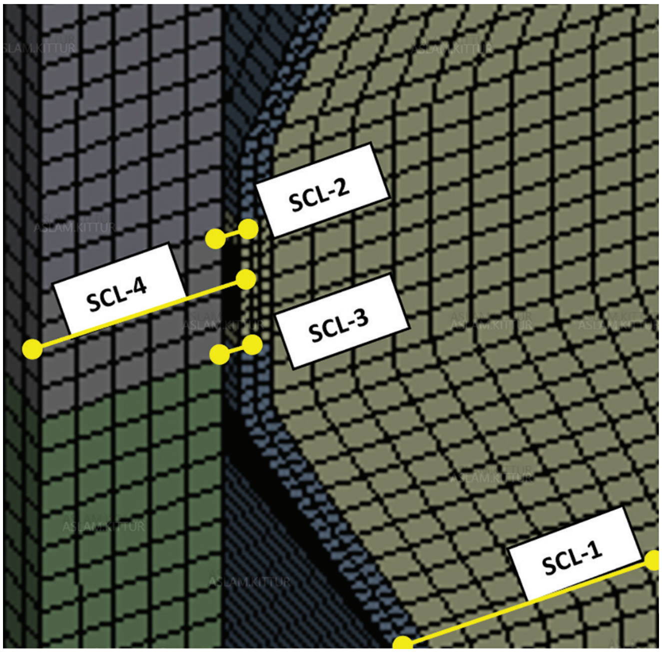

FIG. 7. SCL: Plug weld.

The cases for analysis were developed based on the "objective of assessment" and "methodology" for the below scenarios:

- Scenario 1: Internal design pressure (2,480 psig)
- Scenario 2: Radial thermal gradient of 760°F–720°F = 40°F per 9 in. = 4.45°F/in.
- Scenario 3: Internal maximum operating pressure 1 (2,226 psig)
- Scenario 4: Internal maximum operating pressure 2 (2,276 psig)
- Scenario 5: Internal maximum operating pressure 3 (2,326 psig)

Independent finite element analysis (FEA) simulations were completed for the following developed cases, as shown in **TABLE 1.**

The stress results from Case 1 were utilized to determine the primary membrane (P_m and P_L) and bending stresses (P_b) at the general and local locations for evaluating failure due to plastic collapse.

The stress results (P_m P_L $P_b + Q$) from Cases 2, 3 and 4 were combined with the zero-stress condition existing at equipment shutdown (ambient temperature and atmospheric pressure) to determine the ΔS_n (primary + secondary, equivalent stress range) required for evaluating failure due to cyclical loading (ratcheting).

The stress results (P_m P_L $P_b + Q + F$) from Cases 2, 3 and 4 were combined with the zero-stress condition existing at equipment shutdown (ambient temperature and atmospheric pressure) to determine the ΔS_p (primary + secondary + peak, equivalent stress range) and the S_{alt}.

The acceptance criteria for the fatigue evaluation were established as per ASME Sec-VIII Div-2. The applicable stress intensity factor was $k = 1$ for the design load condition for the kS_m.

The S_m values were chosen for the shell base metal = A-336 GrF22, as it is the lesser of the two materials considering the weld overlay (WOL):

- At a design temperature of 800°F, the S_m = 22.9 ksi.
- At an operating temperature of 740°F, the S_m = 23.3 ksi.

TABLE 1. Load cases for FEA simulation

Cases	Simulation	Shell internal pressure	Shell thick radial thermal gradient	Scenario combination
Case 1	Design	2,480 psig	Not applicable	Scenario 1
Case 2	Maximum operating 1	2,2226 psig	4.45 °F/in.	Scenarios 2 + 3
Case 3	Maximum operating 2	2,276 psig	4.45 °F/in.	Scenarios 2 + 4
Case 4	Maximum operating 3	2,326 psig	4.45 °F/in.	Scenarios 2 + 5

The S_m values for the shell internals A312 Gr321 are:

- At a design temperature of 800°F, the S_m = 16.9 ksi.
- At an operating temperature of 740°F, the S_m = 17.2 ksi.

The four basic stress intensity limits were satisfied as:

1. The allowable value of P_m produced by design internal pressure and other specified mechanical loads—but excluding all secondary and peak stresses—will be less than S_m.
2. The allowable value of P_L produced by design internal pressure and other specified mechanical loads—but excluding all secondary and peak stresses—will be less than $1.5S_m$.
3. The allowable value of ($P_L + P_b$) produced by design pressure and other specified mechanical loads—but excluding all secondary and peak stresses—will be less than $1.5kS_m$.
4. The allowable value of ($P_L + P_b + Q$) produced by design pressure and other specified mechanical loads—but excluding all secondary and peak stresses—will be less than $3.0kS_m$.

Results: Elastic stress analysis. The values of the primary local (membrane + bending) stresses ($P_L + P_b$) at SCL-4 in

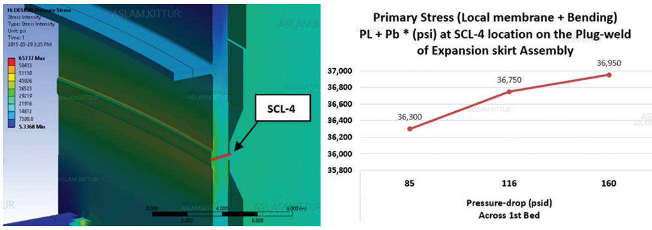

FIG. 8. Expansion skirt plug weld.

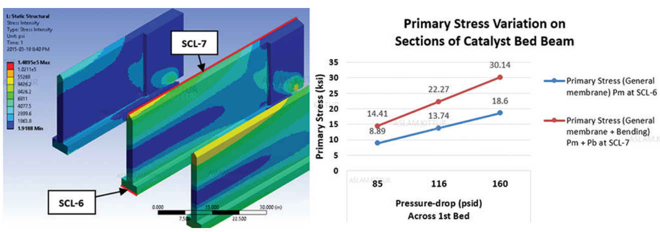

FIG. 9. Catalyst support t-beam.

the plug weld of the expansion skirt rings welded to the reactor shell (**FIG. 8**) were found to be slightly above the maximum allowable of $1.5S_m$.

The maximum stress values obtained for the reactor internal structural components were acceptable and well within the design limits, corresponding to an 85-psid pressure drop across the first bed.

Further simulations for the Case 3 (116 psid) and Case 4 (159 psid) values of pressure drops across the catalyst structure and the expansion skirt structure of the reactor first bed were completed.

It was found that the stress $(P_L + P_b)$ at the expansion skirt welded connection (SCL-4) location does not increase beyond 37 ksi even with an increased value of the pressure drop, which does not correlate to a primary load. It was found that the pressure drop load does not have a direct effect on values of the primary local (membrane + bending) stresses at SCL-4. Based on this finding, an important aspect was realized that the bending stress at SCL-4 in reality should be classified as "secondary" (Q), although membrane stress is classified as "primary" (P_L). As a result, evaluating $(P + Q)$ at SCL-4 at its design-value load is no longer valid with a lower allowable stress limit of $1.5S_m$. In the case of SCL-4, to evaluate the $(P + Q)$ criteria, the more appropriate operating value loads were applied and a corresponding higher allowable stress limit of 3.0Sm was utilized. The maximum simulated stress values of the respective SCL-4 location were found to be well within the allowable limit of $3.0S_m$.

For SCL-6 and SCL-7 in the longest catalyst support beam, as shown in **FIG. 9,** the stresses increased to a level that exceeded the corresponding maximum allowable stress intensity value of 16.9 ksi (S_m for primary general-membrane) and 25.23 ksi ($1.5S_m$ for primary general-membrane + bending).

Takeaway. For the expansion skirt weld, at SCL-4 location on the internal expansion skirt weld as shown in **FIGS. 8** and **10,** it was found that the 9.5-in. thick reactor shell undergoes an insignificant amount of hoops deformation due to the internal pressure of 2,480 psi. The 1-in. thick internal expansion skirt rings are not designed for hoops stress, as the reactor shell internal pressure acts hydrostatically on all internal surfaces of expansion skirt rings. Owing to the weld between the internal expansion skirt rings and the shell, the shell imposes its hoop's deformation on the internal expansion skirt rings at their connection (SCL-4). Although this hoop's deformation is insignificant for the 9.5-in. thick reactor shell, but it becomes significant for the 1-in. thick internal expansion skirt ring, causing the internal expansion skirt ring to preferentially bulge at the welded connection (SCL-4) location, resulting in the above 36 ksi.

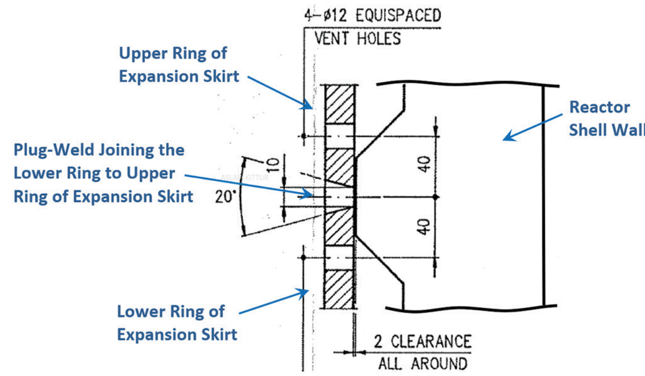

FIG. 10. Plug weld: Expansion skirt assembly.

It has been found that the pressure drop load does not have a direct effect on values of the primary local (membrane + bending) stresses at SCL-4, but the stresses at the SCL-4 location are related to *only* the hoop's deformation imposed by the shell on the internal expansion skirt rings at their connection (SCL-4). Therefore, the membrane + bending stresses are purely "secondary" (Q) in nature. In conclusion, the stress situation of 36 ksi will continue to exist at this magnitude with the presence of internal design pressure, irrespective of change in the magnitude of the pressure drop load.

For the catalyst support beam, an additional simulation was performed with a reduced pressure drop of 142 psid across the first bed, which resulted in a limiting value of the linearized stress intensity of 16.8 ksi at SCL-7 (in the direction of the beam axis) location (**FIG. 9**).

The analysis established that the longest catalyst support beam limits the pressure drop across the first bed of the reactor vessel to 140 psid. For operation of the hydrocracker reactor vessel, this pressure drop could be considered as the limiting value. HP

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[1] American Society of Mechanical Engineers (ASME), "Boiler and pressure vessel code ASME Sec-VIII Div-2.
[2] American Society of Mechanical Engineers (ASME), "Boiler and pressure vessel code: Material properties," ASME Sec-II.

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The phosphorus trap is renewed

One of today's hottest industry topics is renewable fuels. Driven by the desire for a sustainable future, governments are enacting legislation to reduce greenhouse gas emissions. In the U.S., Renewable Fuel Standard 2 (RSF 2) requires refineries to blend 36 Bgal of renewable fuel into the country's total transportation fuel consumption by 2022. In Europe, Renewable Energy Directive II (RED II) requires all EU countries to ensure that at least 14% of their transport fuels come from renewable sources by 2030.

A few companies are now starting to embark on technology offerings for this growing market segment. The four key elements to consider when commissioning a renewable fuels unit are:

- Feed sourcing
- Feed pretreatment
- Hydroprocessing
- Dewaxing.

The author's company began researching renewable fuels production approximately 20 yr ago and started up the first unit with its licensed technology for hydrotreating renewable feedstocks[a] utilizing its proprietary catalyst in 2010. With involvement in more than 60 renewable fuels plants, the licensed technology[a] and more than 125 yr of operating experience, the company is one of the foremost experts—and one of the most successful licensors—for renewable fuels needs. The company provides chemical and fuel producers with solutions for the world's most energy-efficient production, resulting in a lower carbon intensity (CI) score. In addition, its experience includes the full gamut of feedstocks used to make renewable fuels. The combination of this industrial experience and a commitment to research and development has led to the introduction of a renewable hydrotreating catalyst[b].

Feedstock impurities. Various types of renewable feedstocks are available to produce transportation fuels, including:

- Oilseed crops (e.g., soybean oil, canola)
- Tall oil, corn oil, used cooking oils and animal fats
- Lignocellulosic biomass from agricultural residues, algae, trees and grasses.

Compared to the processing of fossil fuels, the conversion of renewables to transport fuels introduces new types of contaminants, such as phosphorus, potassium, sodium, calcium, magnesium, arsenic and more. These contaminants come from living tissues, such as cell membranes, bone dust, muscle residue and contaminants introduced in processing. Phosphorus is the most common: phospholipids are the primary building blocks of cell membranes and are present as inorganic phosphorus in bone dust. While the level of phosphorus is different for each renewable feedstock, and different pre-treatment technologies are available to reduce such impurities, some amount is always present entering the hydroprocessing reactor. These impurities can lead to rapid pressure drop build-up if the appropriate grading/trap material is not installed in the reactor.

Phosphorus: Understanding pressure drop mechanism. The author's company is always conducting fundamental research to understand the mechanisms at play, whether for reaction kinetics or pressure drop build. Using a state-of-the-art scanning electron microscope, it could discern the reason phosphorus led to high-pressure drop build in renewable units using conventional grading products.

FIG. 1 shows a traditional particle used for trapping contaminants, and it is clear that the phosphorus does not penetrate the pore system but rather binds itself to the catalyst surface, forming a type of glass material. This glass-like material glues the catalyst particles together, filling up the interstitial void and resulting in rapid pressure drop build. Zooming in on the catalyst surface (**FIG. 2**) shows the high phosphorus (P-) containing crust around the catalyst.

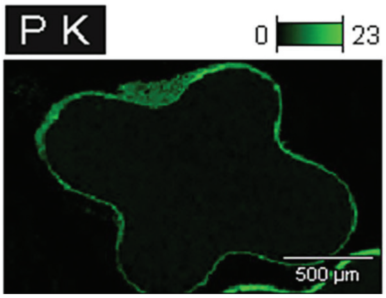

FIG. 1. Catalyst particle showing phosphorus sitting on the catalyst surface. Image produced using an energy dispersive spectrometer attached to a scanning electron microscope.

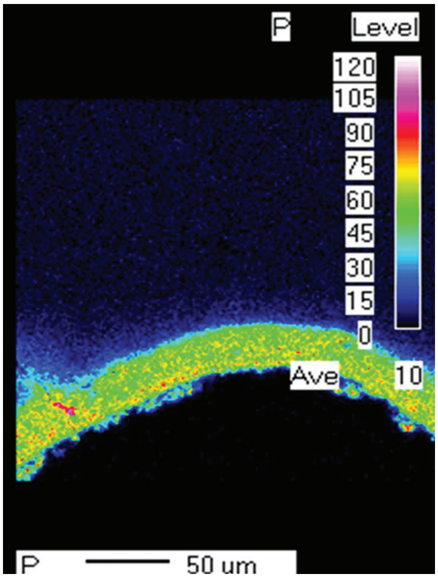

FIG. 2. Crust with high phosphorus content.

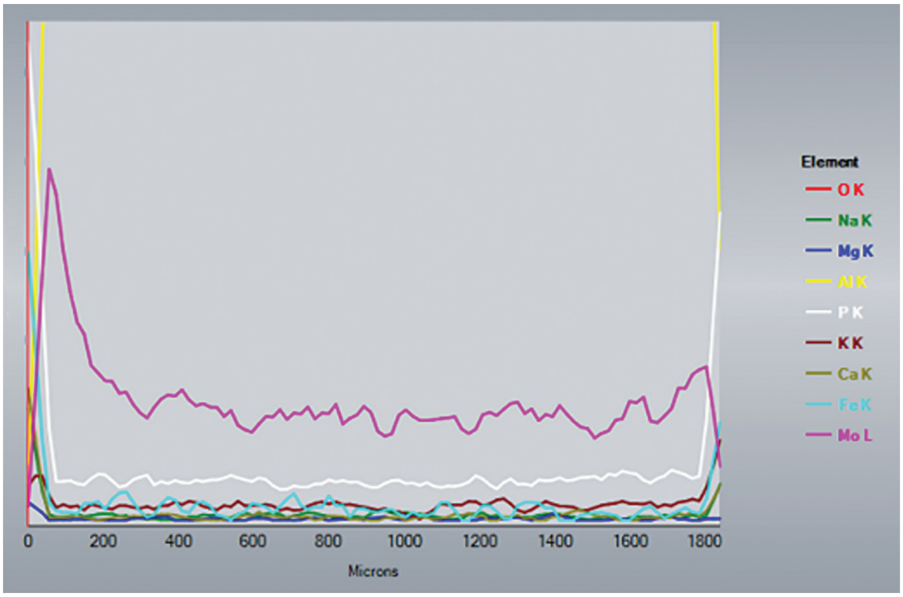

FIG. 3. Phosphorus profile from first-generation phosphorous traps.

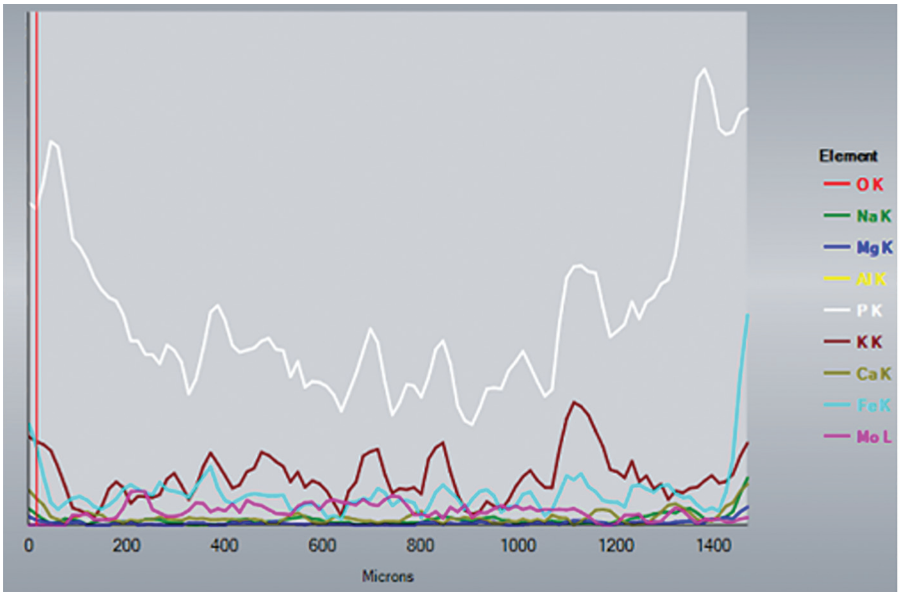

FIG. 5. Phosphorus profile of renewable hydrotreating catalyst[b].

Original improvement of existing products gave better phosphorus uptake and penetration but did not completely solve the fundamental issue of phosphorus covering the surface (**FIG. 3**).

Using this fundamental understanding of both the crust formation mechanism and the nature of the phosphorus-containing molecules, plus applying knowledge of catalyst design and manufacturing, led to the development of the renewable hydrotreating catalyst[b] developed for fixed-bed HDO service applied as a guard in diesel and jet hydrotreating units processing renewable feedstocks. The large-pore catalyst with a unique pore system is designed to maximize phosphorus pick-up from renewable feedstocks and is characterized by high capacity and low HDO activity.

It exhibits full penetration into the pore system (**FIG. 4**). Additionally, the profile shows much higher levels of phosphorus pick-up in the center of the particle and higher overall area under the curve, which translates to overall higher capacity (**FIG. 5**).

This translates to three to four times (and in some cases even six times) higher phosphorus pick-up. The first load to include the proprietary catalyst[b] was installed in 2020 and has subsequently been installed in a handful of units, performing well in all instances. The main advantages are that it:

- Prevents phosphorus slip to the bulk catalyst
- Inhibits pressure drop build from phosphorus crust formation
- Ensures longer catalyst cycle length, improving unit profitability and catalyst value ratio.

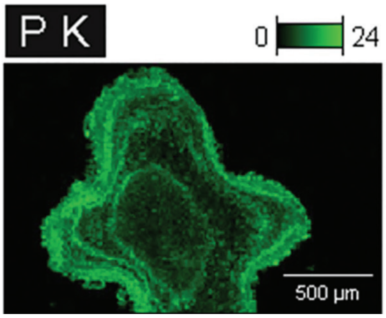

FIG. 4. Catalyst particle showing phosphorus penetrating into the pore system of the renewable hydrotreating catalyst[b].

Takeaway. As always, getting the best value from a grading and catalyst system is two-fold: the technology itself; and (just as, if not more important) the technical expertise needed to deliver the most value. The proprietary catalyst[b] will help extend the life of hydroprocessing catalysts in a standalone or co-processing renewable fuels unit. Chemistry is behind almost any product or fuel that defines the way we live today, and it will be just as important in shaping the future. Reducing carbon emissions of fossil fuels together with production of renewable fuels will be part of the future energy transition. HP

NOTES

[a] Haldor Topsoe's HydroFlex unit
[b] Haldor Topsoe's TK-3000 PhosTrap™ (patent pending)

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Using ANN modeling to predict the melt flow index in real time

Controlling polymer quality during manufacturing is a significant challenge due to the cost, maintenance and reliability of online measurements of final product quality. Polymerization is highly nonlinear in nature, which poses great challenges for manufacturers to produce consistent quality product. In most cases, the quality measurement is done manually through sample lab analysis in post-production.

These challenges call for a real-time solution. Most manufacturing companies are adopting advanced digital technologies to optimize operations. Several open-source programming tools are available, along with machine learning (ML) and artificial intelligence (AI) algorithms and high-performance computing infrastructures. These advancements are helping manufacturers to improve their operational performance in real time.

Using available historical process data, this article details how an artificial neural network (ANN) model was developed and deployed for estimating the melt flow index (MFI)—an important polymer property of the polyethylene (PE) production process—in real time. The model's results were validated using training, testing and real-time data. Prediction was found to be within the expected accuracy for multiple grades—even during the grade transition period. This finding facilitated optimizing the production process and expensive laboratory resources. In addition, it helped to sustain operations despite limited laboratory human resources due to the COVID-19 pandemic.

The MFI is an extremely important parameter in all stages of the polymer lifecycle, from polymer synthesis/manufacturing and polymer processing to final product performances. The MFI is correlated to many of the polymer product properties, such as tensile strength, rigidity/stiffness, hardness and impact strength.[1] A low-density PE (LDPE) plant produces various grades of MFIs; therefore, there is a need for altering the process parameters based on the product quality. Prediction of product quality is a very important tool for steady-state operation and during grade transitioning.

The MFI of polymers is measured as the weight of the polymer (in grams) flowing in 10 min through an instrument of a specific diameter and length by applying a certain pressure and temperature. It is evaluated in a time-consuming, worker-intensive and expensive laboratory analysis. Available online instruments are also very expensive and require a high maintenance cost, with reliability not a certainty. Due to the unavailability of real-time MFIs, operators run reactors based on their experience and judgment, as well as on laboratory analysis feedback, which has a time lag due to the process holds and sampling time required in the laboratory. This delay may result in the generation of higher off-specification material due to delayed operating actions. Ultimately, this can lead to a loss of margin, since the downgraded product commands a lower price in the market.

The MFI is influenced by many process variables, such as temperature, pressure, catalyst and monomer rates, as well as additive flowrates and their concentrations. These variables are commonly measured in real time. With the availability of a large amount of data, a high-speed computational facility and AI algorithms (e.g., ANN), it is possible to predict the MFI in real time. Jumari and Yusof compared the performances of a first-principles model and an ANN-based model on MFI predictions in industrial polypropylene (PP) loop reactors and found that the ANN model was more accurate than the first-principles model.[2] As the ANN model required less central processing unit time, it could run in real time, which was a challenge for the first-principles model. Liu, Shi and Sun, within their work on predicting the MFI of PP processes by using an ANN model, concluded that predictions using the ANN model were reliable and accurate for practical use.[3]

In this article, an ANN-based model was developed, and then validated with the historical data and tuned to obtain the required accuracy. After this, the model

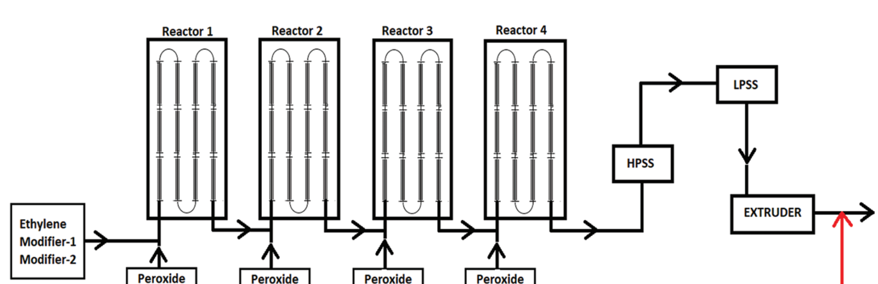

FIG. 1. Simplified process block diagram of the LDPE plant.

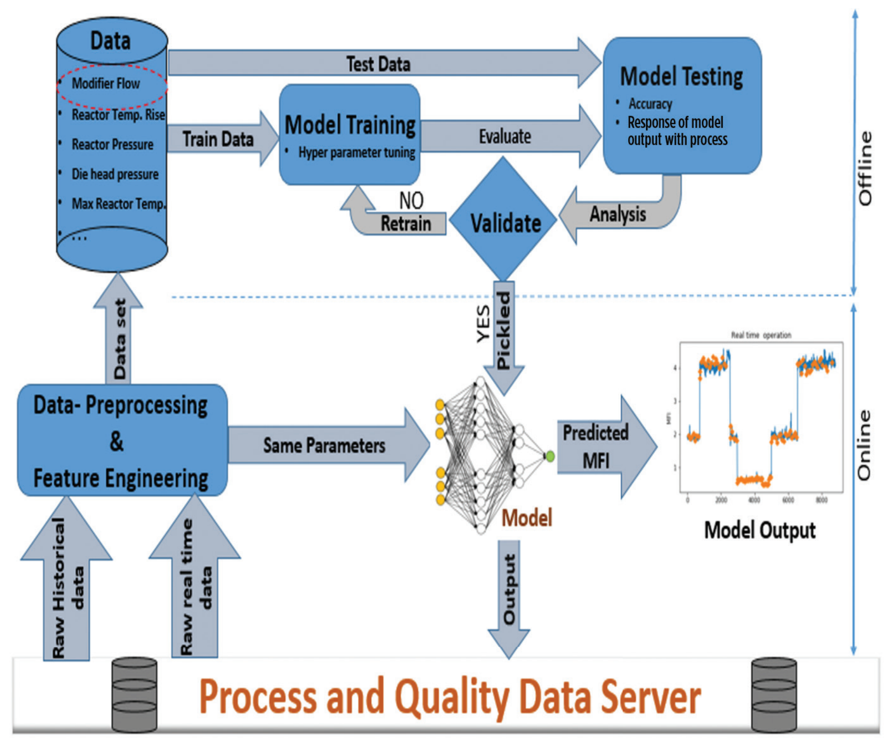

FIG. 2. ANN model development, validation and deployment workflow.

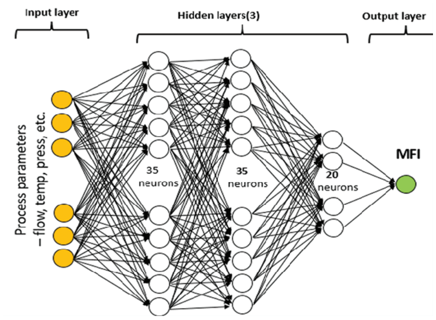

FIG. 3. ANN model structure.

was deployed online, and prediction results were validated over a period and over various grades. It was observed that the model's results and responses conformed to actual values within a defined accuracy range. The model also responded to the changes in process conditions in desired directions. The response during the grade transition was also acceptable. The model's predictions were used for plant operations, thus reducing the need for expensive laboratory analysis.

Process description. The ANN model was used in an LDPE plant, where the LDPE was produced in a very high-pressure tubular reactor. A simplified version of the existing plant is shown in **FIG. 1.** The process gas and the mixture of ethylene and modifiers were heated in the preheater before entering the reaction zone. The complete reaction occurred across four reactors in a series. A free radical initiator was pumped into the four reactors to start the polymerization reaction. Multiple modifiers in different proportions were added to vary the MFI. Polymerization was a strongly exothermic reaction. Unreacted reactants were separated in high- and low-pressure separators before sending the polymer to the extruder, after which it was bagged in the packing section. Multiple grades of LDPE were produced and characterized primarily by MFI/density.

The ANN model. The workflow for the model development, validation and deployment of the ANN model are shown in **FIG. 2.** All raw historical data was taken from process and quality data servers. For the development phase, the raw data underwent pre-processing and feature engineering to finalize the dataset for model input. The model was then trained, using a training dataset, and validated against test data. Upon successful validation, the model was then picked for real-time prediction. Real-time data also came from the same server and underwent the same pre-processing and feature engineering before being inputted into the model. The online deployed model was scheduled to run at a particular frequency (5 min in this case), as it provided enough time for the operator to respond to the changes in the MFI.

Data pre-processing, feature engineering and transformation. In any polyolefin reaction, the kinetics of the reaction are very important. It is also important to take care of process lags at the time of dataset preparation. In this case, while MFI sampling was done from the extruder end, the reaction occurred across four reactors, followed by the product passing through a high-pressure separator section (HPSS) and a low-pressure separator section (LPSS) to reach the extruder. Time lag was considered for the proper match of the target variable to the process parameters. Interaction analysis of the process variable to the MFI indicated that it took approximately 20 min for any changes done on the inputs to be reflected in the MFI. Out of these 20 min, materials remained in the reactors for about 10 min, while the remaining time was consumed to pass through the HPSS and LPSS. To provide the impact of the transportation lag while making input data sets, a lag of 15 min was considered for parameters until the second reactor, 10 min for the third and fourth reactor parameters, while zero lag was considered for the extruder parameter.

On understanding the process and interaction of the variables to the MFI, 21 variables were identified. These included flow (such as fresh feed, recycled flow, modifiers and free radical initiators at different reactor locations), temperature (all four reactor inlets, and maximum and end temperatures) and pressures (reactors, extruders and feed, among others). With more than 21 parameters and time lag, a dataset was prepared with 2 yr of historical data. After outlier treatment, the correlation was checked for each input parameter with the MFI.

On analysis, it was found that the initiator flowmeter did not measure pure initiator content, as this measurement was done after the addition of solvent. This called for additional parameter generation through feature engineering to map the proper response to the MFI. The initiator impacted the extent of the reaction, which led to a temperature increase across the reactor. This increase also indirectly indicated the same extent of the reaction. Therefore, in the following, the rise in temperature

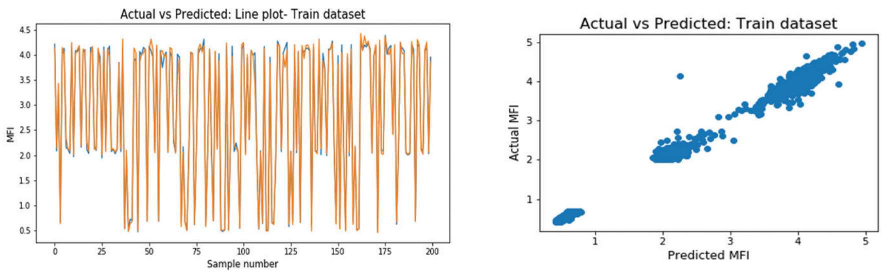

FIG. 4. Offline validation: Training dataset.

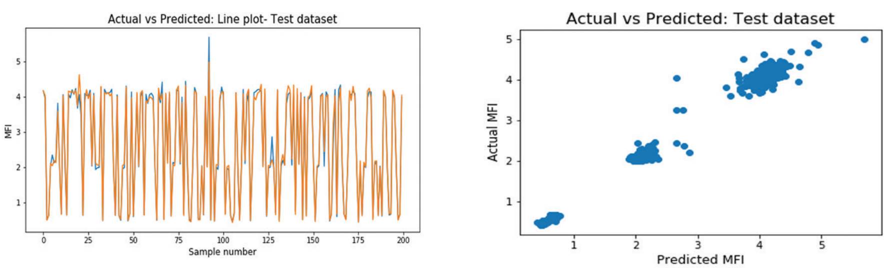

FIG. 5. Offline validation: Testing dataset.

across each reactor was considered, instead of the solvent mixed initiator flow:

- Reactor 1 temperature rise = (RX-1 maximum temperature) – (preheater A/B O/L temperature)
- Reactor 2 temperature rise = (RX-2 maximum temperature) – (RX-1 end temperature)
- Reactor 3 temperature rise = (RX-3 maximum temperature) – (RX-2 end temperature)
- Reactor 4 temperature rise = (RX-4 maximum temperature) – (RX-3 end temperature).

To normalize the modifier flow with the monomer flow rate, a ratio of modifiers was included and individual parameters were dropped:

- Modifier 1 flow/C_2 flow = Modifier 1 flow/net C_2 flow
- Modifier 2 flow/C_2 flow = Modifier 2 flow/net C_2 flow

where:

Net C_2 flow is calculated from the following:

Fresh ethylene flow + low-pressure recycle flow – purge gas flow.

The final dataset contained Modifier 1 flow/C_2 flow; Modifier 2 flow/C_2 flow; reactor pressure; Reactor 1 temperature rise; Reactor 1 maximum temperature; Reactor 2 temperature rise; Reactor 2 maximum temperature; Reactor 3 temperature rise; Reactor 3 maximum temperature; Reactor 4 temperature rise; Reactor 4 maximum temperature; Reactor 4 end temperature; die head pressure of the extruder; and the MFI (the target of the model), which was used for modeling. There were 13 predictors and one output.

ANN model tuning. The model architecture, including hyper parameters, is briefly described here. The input data set was split into an 80-20 ratio for training and testing purposes. Training and testing data were then fitted and transformed using the standard scalar method. The transformed data was fed as input into the ANN model (**FIG. 3**). The model was fine-tuned with different hyper parameters, which included the hidden layers, the number of neurons, the activation function, the optimizer, and the regularization to obtain the desired accuracy and acceptability.

The fine-tuned model had the following structure:

- Number of input variables: 13
- Number of hidden layers: 3
- Number of output variables: 1
- Number of neurons in each hidden layer: 35, 35 and 20.
- Activation function: The rectified linear unit (ReLu) and optimizer used a limited-memory Broyden-Fletcher-Goldfarb-Shanno (BFGS) algorithm, with a regularization parameter (alpha) of 1.41.

Offline validation. The metrics used for validation were root mean square error (RMSE) and the percentage of data points

TABLE 1. Offline validation statistics

MFI range		RMSE	Data points within R&R, %
0.5	Train	0.0335	90.05
	Test	0.0331	90.38
2.5	Train	0.074	96.09
	Test	0.091	92.48
4.5	Train	0.113	96.73
	Test	0.155	92.72

TABLE 2. Real-time validation statistics

MFI range	RMSE	Data points within R&R, %	No. of samples
0.5	0.0379	91.66	36
2.5	0.0751	96.29	27
4.5	.0133	97.2	36

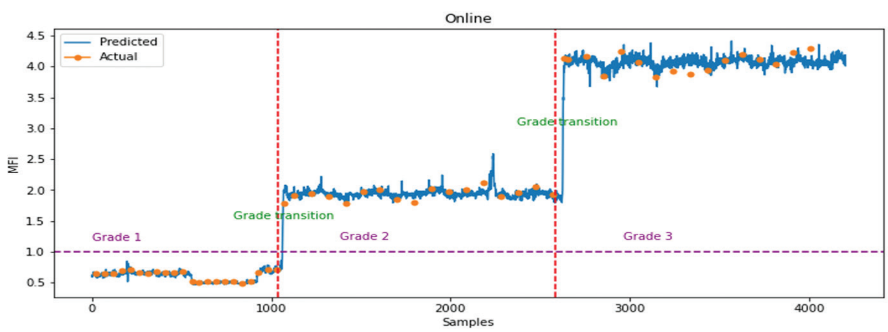

FIG. 6. Online validation with grade changeover.

within gage repeatability and reproducibility (R&R). Typically, when an instrument is used for measuring any physical values, there are some inherent errors in measurement due to the instrument. Repeatability is the variation in measurements by a single instrument or person under the same conditions—whereas, reproducibility is the variation in measurements when the experiment is reproduced in an entirely different condition/lab. As per the accepted criteria, the authors have considered that the error should be within R&R at least 90% of the time.

Trends during training and testing were also observed during the validation analysis. **FIGS. 4** and **5** correspond to the training and testing trends. Testing trends showed that the prediction was in conformance with the actual value. **TABLE 1** provides the error analysis. To avoid the clutter, only a section of predicted and actual values was plotted in the line chart, whereas the scatter chart shows the whole dataset.

Online validation. After satisfactory offline validation, the model was deployed online for real-time validation. **FIG. 6** shows the online trend, while **TABLE 2** provides the error analysis for real-time performance.

Model responses were also validated to see if the model was responding correctly with the changes in process variables in positive and negative directions. **FIG. 7** shows the responses of model prediction for changes in one of the important variables. With an increase in modifier flow, the MFI should increase, which was correctly reflected by the model. Similar validation was done to check the correctness of the model response with the directional changes of other important process variables. The correct and accurate responses provided the operating team with the confidence to adopt the model results for optimum reactor operation.

Takeaway. The model results agreed with the measurement R&R, while the model was kept for online validation for approximately 3 mos. After successful validation, the ANN model results were used by the operator for reactor operation, enabling

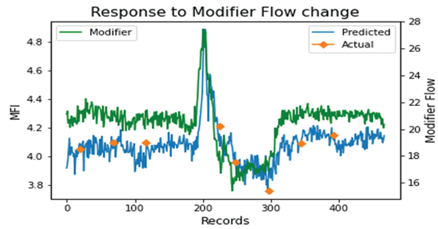

FIG. 7. Model response with modifier flow change.

lab measurements to be reduced by around 50%. Operations are now responding to product quality changes more dynamically than before, enabling optimum reactor operation and reducing grade demotion, thereby improving margin. HP

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Impact plant performance by improving the steam system

Petrochemical plants and refineries face production and profitability challenges on multiple fronts. Some of the typical difficulties encountered include:

- An unnecessarily poor energy intensity index or a low-energy efficiency index
- Unplanned maintenance expenditures to fix damaged and unreliable steam-using equipment (e.g., turbines, flares, reboilers and ejectors)
- Wasted operating expenditures and experiencing product loss on avoidable items (e.g., turbine trips, product lines freezing, reboiler control issues or equipment failures)
- Hassles working around process problems, such as production bottlenecks, low cutpoints and degrading vacuum levels
- Concerns on safety and environmental pollution
- Lack of sufficient steam specialist knowledge to address problems and opportunities
- Trouble applying limited resources to make improvements when the priority is to keep the plant running within budget
- Difficulties accurately predicting and demonstrating the true savings of improvements.

Plant operations are commonly focused on production, and so they tend to accept energy losses and poor reliability as accepted practices. As a result, substantial economic opportunities become invisible. Plants may not know where to start to improve in areas such as safety, environmental impact, production rates, product quality, equipment reliability and/or energy efficiency. One way of impacting all these areas is to treat the plant's steam system as a crucial heat asset, recognize the importance of the steam trap population to its health, and to implement a simple four-step program to protect and improve it.

Four effective steps that impact plant performance. The following steps are important for improving a plant's steam heat asset:[1]

1. Select the right steam trap for the application
2. Install the steam trap correctly
3. Implement a sustainable steam trap management program
4. Optimize the performance of steam-using equipment.

Select the right steam trap for the application. To properly select a steam trap, it is important to understand:

- The purpose of a steam trap
- How normal operation and failure may affect the purpose of a trap
- How generic steam trap types operate
- How steam traps can fail (common failure modes)
- Special requirements for specific steam trap applications that should be considered when selecting steam traps
- A site's experience and bias regarding steam trap performance
- Factors affecting lifecycle costs (for reliability and total cost of ownership)
- Specific performance criteria for individual steam trap models
- How to accurately model steam trap lifecycle costs.

Application-specific guidance. One of the most critical areas for steam trap selection is considering the application to account for specific challenges. The most common steam applications in hydrocarbon processing facilities that require careful trap selection include steam distribution lines, steam turbines, process heaters, flares, sulfur pit coils and steam tracing lines (particularly in the transport of high-viscosity products like sulfur or resid).[2]

Steam distribution. Steam is generated—as either superheated or wet quality—and then distributed throughout the facility to provide process heat to equipment.[3] It is essential that condensate forming in the steam distribution system is drained immediately and not allowed to back up, as this could result in water hammer damage and the reduction of process equipment reliability.[4] In general, for these applications, steam traps with near-instantaneous discharge provide the best performance. There is commonly the need for these traps to vent air, so the selection process should consider all requirements. Subsequently, float and thermostatic traps, as well as thermodynamic traps containing thermostatic air vents, and thermostatic balanced-pressure capsules, are recommended.[5] Bimetal traps should never be used for draining steam mains due to their operating principle requiring deep levels of subcooling, which can lead to condensate backup. There may also be some special requirements for steam main drainage (e.g., to drain the boiler header piping, which follows immediately after steam boilers/steam generators, for high-temperature super-heated steam lines or where traps are widely spaced).

Steam tracing. Steam tracing typically accounts for most steam trap installations in a hydrocarbon processing facility.

One key purpose of steam tracing is to maintain temperatures of process fluids and, in critical cases, to keep product in a process fluid transfer line from freezing (or solidifying). The heat load is typically about the same as the radiant, convective and conductive heat loss from the line. As lines are usually well insulated, the heat loss and resulting condensate loads may be minimal. This may present challenges and also opportunities for steam trap selection, particularly when there are loops or lifts that flow steam and condensate upward.

Steam tracing can generally be categorized as high temperature or low temperature. In critical high-temperature applications, condensate backup is undesirable. A steam trap with an instantaneous and continuous discharge with a tight shutoff may be preferred for maintaining the lowest lifecycle costs, and thermodynamic disc traps or balanced-pressure thermostatic traps may be acceptable for lowest-initial-cost installations. Copper tracing lines experience corrosion over time and may, therefore, present plugging difficulties. This scenario is best mitigated by using a thermodynamic trap with enlarged ports and short passages, or, for low-temperature applications, a bimetallic steam trap. Bimetal traps with a low set temperature can also be used for instrument enclosures to mitigate damage to the instrument by excess temperatures.

Steam turbines. Among the most common steam equipment applications in a petrochemical facility are steam turbines providing power to pumps, compressors or power generators. Turbines are very sensitive to condensate in the steam supply and to condensate backup in the governor casing, body or exhaust lines. Consequently, the best choice for this application is a steam trap with instantaneous and continuous discharge, such as a float and thermostatic design. However, thermodynamic disc traps may be used with appropriate drainage pockets and drain lines. Bimetal steam traps should never be used because of the possibility of condensate backup or steam loss.

Turbine condensate drainage should be carefully considered to ensure that all appropriate points on the upstream steam supply, governor, casing and exhaust lines are all properly drained.[4] A slug-tolerant separator should be considered on the steam supply line immediately before critical turbines. Traps on the exhaust should be appropriately sized for the turbine's efficiency and operation, considering that the exhaust steam may be significantly wet or may contain a substantial condensate load.

Process heating equipment. Process heating equipment may include applications such as reboilers, heat exchangers, air heaters, evaporators, concentrators, dryers and aboveground storage tank coils. These applications may have various control strategies and operating conditions that may affect the range of steam pressures and flowrates. In some cases, a steam trap may not work correctly due to a stall condition.[6,7] This may occur when the ideal steam pressure in the equipment for steady-

TABLE 1. Recommended trap type according to application—Legend: First choice (1), second choice (2) and third choice (3)

Service/application	Free float and thermostatic	Thermodynamic	Bimetal adjust-temp	Balanced pressure thermostatic	Non-electric pump/trap	Non-electric pump	Level pot with outlet control
				Tracing			
Tracing lines[9]							
Low temperature, < 200°F	–	–	1	2	–	–	–
High temperature, > 200°F	1	3	–	2	–	–	–
Jacketed pipe	1	2	–	–	–	–	–
				Drip			
Saturated steam main: ≥ 20, ≤ 650	1	2	–	–	–	–	–
Superheated steam—all services[6]: ≥ 20, ≤ 650	1	2	–	–	–	–	–
Flare stack steam lines							
Steam to control valve	1	–	–	–	–	–	–
Control valve outlet to flare stack	1	–	–	–	1	–	–
Turbine supply lines	1	2	–	–	–	–	–
Turbine casing drains:[7,8] ≥ 20, ≤ 650	1	–	–	–	–	–	–
Turbine exhaust steam lines: > 0, ≤ 150	1	–	–	–	–	–	–
				Process			
Process heaters: ≤ 200 supply							
Stall condition	2	–	–	–	1	2	–
No stall condition	1	–	–	–	–	–	–

state operation is lower than the discharge pressure after the steam trap. In these conditions, a secondary pressure drainer pump/trap combination may be necessary.

When steam traps are appropriate, the best selection is a float and thermostatic-style trap with instantaneous continuous discharge, incorporating reinforced components to avoid water hammer damage. Disc traps may sometimes be utilized for tank coils, comfort heaters and small noncritical heat exchangers, but they are not suitable for other process heating applications. Inverted buckets may not be suitable because of their cyclical operating characteristics or because of challenges faced with non-condensable air in the system. Bimetal traps are generally not recommended due to their operational reliance on subcooling and their sensitivity to changes in back pressure, which may result in condensate backup. This, in turn, may flood equipment, reducing the heat transfer area.[8] Bimetals may be considered for small storage tank coils, where only approximate temperature regulation is required, and also where it is not practical to install a steam control valve and control system.

Flares. A steam-assisted flare requires a continuous low volume of steam that is often supplied through an orifice plate bypass around control valves to keep the flare tips hot while the flare is not burning vapors. When flare gas is supplied to the flare, the steam control valves open to supply steam for atomization and flame stability. If wet steam is supplied, the flare tips may become eroded. This can reduce the effectiveness of the steam distribution at the tip and may cause unclean "sooty" burning that could result in environmental issues. If the flare steam lines upstream and downstream of the control valves are not properly drained, then water slugs may be propelled to the flare tip. This may severely damage the flare tip or extinguish the pilot light before the flare gas is ignited. Both conditions may result in environmental issues. Repair of the flare tips or complete candelabra replacement may also be costly.

Due to the criticality and design of flare steam lines and control stations, the best choice of steam trap, such as a float and thermostatic style, will instantaneously and continuously discharge condensate. Bimetal traps must not be used. Thermostatic traps and inverted buckets are not a first choice. Disc traps upstream of the control valve station may be acceptable if the condensate collection pockets and the drain lines are properly sized and designed. Special selections may be required for drain locations after the control valve stations, which may have extremely low operating pressures in standby operation and may be close to grade.

Sulfur pit coils. Sulfur pit coils are below ground, and steam traps usually must be installed above ground. Condensate must be lifted from the coil to the trap. This can commonly cause steam locking of the trap, which may prevent condensate discharge and result in lowered temperatures, coil damage and possible solidifications. Operators may then have to open drainage location bypasses and blowdown valves to maintain sulfur temperatures, which can cause energy loss and safety issues. Steam locking of the sulfur pit coil may be mitigated by the installation design and by selecting a trap with a small vent hole to address steam locking.[2]

Structured approach to selection and documentation.

A structured approach is necessary to select the trap, and then to document the selection and communicate it to maintenance personnel, engineering contractors and capital project teams. A trusted steam system specialist can help to guide steam trap selection for any type of application and to provide established procedures for documenting and communicating selection standards. **TABLE 1** shows an extract from a typical selection document reviewing steam trap selections.

Install the steam trap correctly.

Steam traps must be properly installed so they can function correctly. Installation details depend on the application, along with the existing factors in the field and the type of steam trap selected. A steam trap is typically installed as part of a condensate discharge location (CDL), which includes other items such as an inlet isolation valve, a strainer with blowdown, the steam trap, an outlet isolation valve, a check valve, a disassembly component such as a

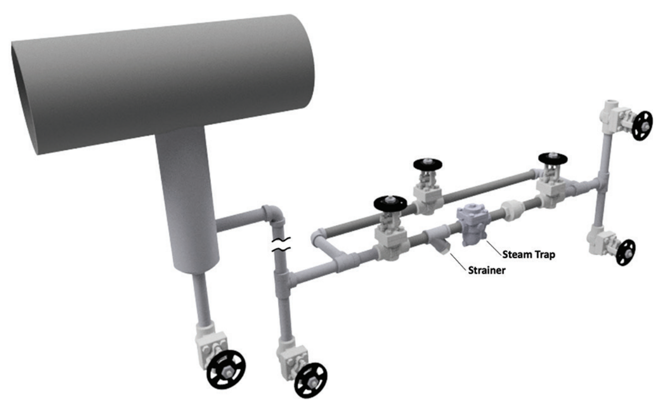

FIG. 1. A properly designed collecting leg/CDL has multiple components that require correct installation.

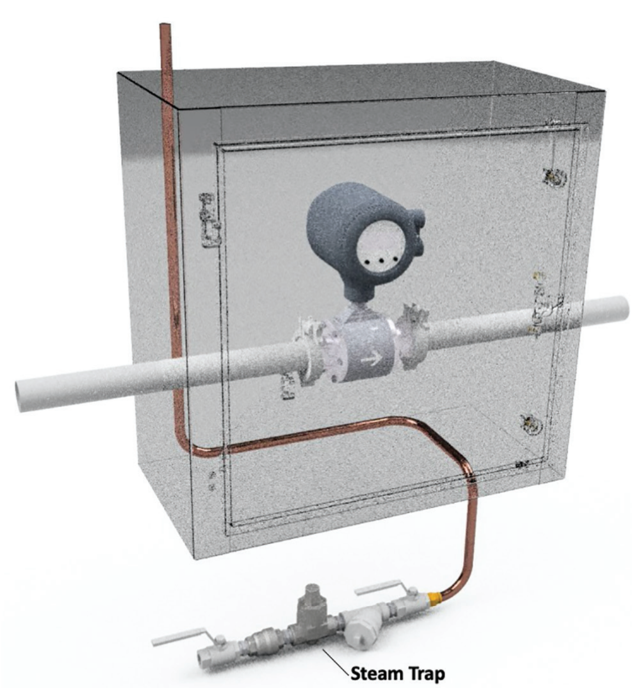

FIG. 2. Instrument enclosures require subcooling traps to avoid burning instruments.

flange or union, valves for depressurizing the steam trap body to a safe zero-energy state to allow for trap replacement, and valves to allow flow to the atmosphere for testing or troubleshooting. The configuration of all these elements critically impacts the performance of the steam trap. **FIGS. 1–6** show typical arrangements for some of the more common types of condensate drainage locations.

For pressures up to 650 psig, the piping and individual components for a CDL may be replaced by a single steam trap station to reduce installation space, complexity, labor and leak points. In some situations, several condensate discharge locations may be in proximity, and there may be a requirement to return condensate. In these cases, it may be appropriate to utilize steam supply manifolds and condensate collection manifolds with trap stations (**FIG. 6**).

It is essential to generate a documented standard for installing steam traps that can be used to:

- Define the trap selection guide
- Provide data to office-based personnel (e.g., planners, procurement personnel and project engineers)
- Convey selection and installation information to capital project groups or to external engineering, procurement and construction (EPC) companies
- Train maintenance and operations personnel.

It is also helpful to generate the same information in the form of laminated posters for maintenance workshops and as pocket-sized flip books.

Implement a sustainable steam trap management program. Even when a steam trap is correctly selected and installed, it will have a finite life. A sustainable steam trap management program is essential to maintain an acceptable failure rate and to monitor the current state of failures and the number of good in-service drainage locations.[9] An effective trap management program should cover the following:

- Accurately and consistently identify steam trap failures; these actions can have a critical impact on program costs, and are especially essential when multiple sites are being compared
- Track survey progress to ensure that all drainage locations are inspected
- Record failure and maintenance data
- Track replacement/maintenance actions

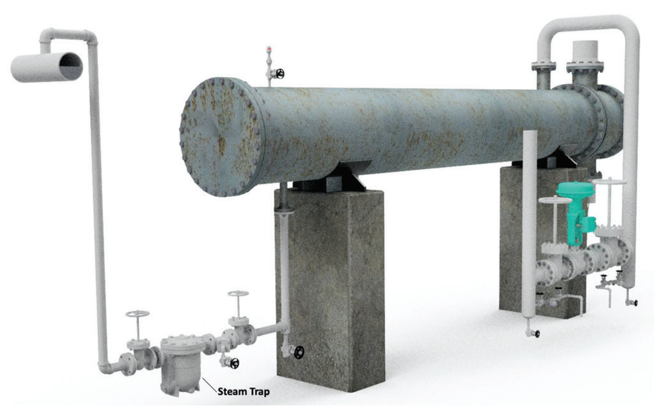

FIG. 3. A heat exchanger sustaining positive pressure differential from equipment to condensate return header may only require a trap for condensate.

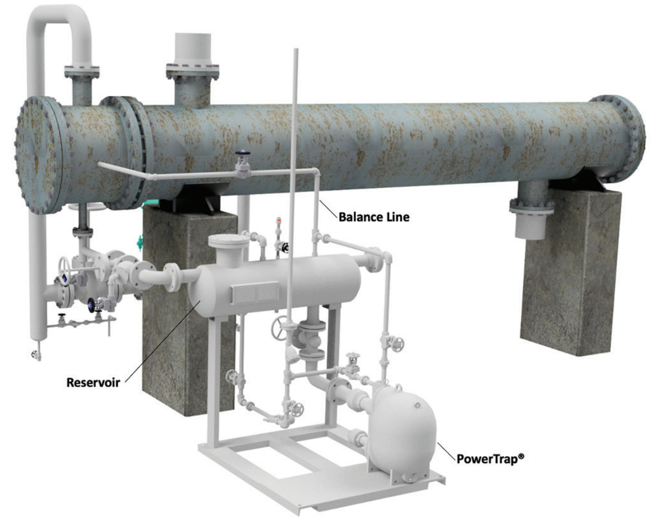

FIG. 4. A heat exchanger experiencing stall conditions may require a combination pump/trap system for condensate discharge.

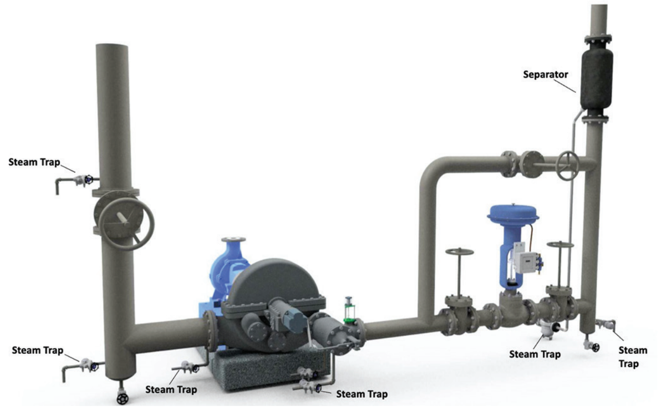

FIG. 5. Turbines may only have a few traps (or none), and this can lead to premature failure.

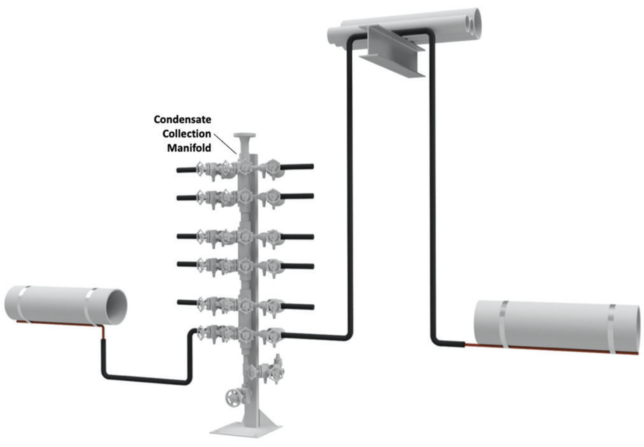

FIG. 6. Tracing systems may require special trap options to mitigate steam locking from uplifts.

- Use recorded data to identify bad actors and root causes of failures
- Generate information on savings to justify expenditures and return on investment (ROI)
- Ensure that the program continues to run smoothly and make continuous improvements year after year.

Optimize steam-using equipment performance. Further steam system asset improvements can be made once the steam trap population is under control, after high-quality steam is supplied to production and effective condensate drainage for steam-using equipment is deployed throughout the facility. The investigative steps for this typically include:

1. Conduct a one-day plant walkthrough and a consultation session with a steam specialist to identify opportunities and potential value. The plant can assess priority and potential ROIs to decide which options to progress. Sometimes, a plant may already have specific challenges for the event to review.
2. Obtain detailed technical assessments of a specific opportunity or challenge onsite to identify the root cause of the issue and collect data to engineer a solution and more accurately calculate potential benefits. This action is usually followed by offsite work to design a solution. Several technical assessments may be combined to create a larger event.
3. Work with plant personnel to complete an implementation proposal to gain funding.
4. Provide equipment, engineering, construction and commissioning support for the solution.

High-value targets for steam-using applications in hydrocarbon processing facilities typically include:

- Steam turbines to reduce trips, increase reliability, decrease damage incidents and reduce plant steam generation
- Reboilers to decrease plant steam generation, reduce failures (gaskets, corrosion of tube bundles and piping erosion), and increase performance to improve production rates and product quality by fixing operating problems, reconfiguring processes to use lower-pressure steam that may be vented and by recovering condensate
- Flares to reduce risk of flare tip damage from erosion or water slugs, and to decrease condensate backflow down the flare gas line
- Condensate recovery strategies to save energy, alleviate raw water treatment limits and reduce environmental impact.

Takeaway. Hydrocarbon processing plants are facing new challenges that are frequently changing, so the focus of these plants is often on the production process and keeping things running. Taking a strategic look at the steam system, treating it as an asset and following a simple four-step program to improve it can positively impact the plants' performance in many areas, including safety, environmental impact, production rates, product quality, equipment reliability and energy efficiency. This article has outlined the four steps to steam system improvement and provides references that provide more detail on implementation. HP

ACKNOWLEDGMENTS

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Design and scale-up of gaseous Group A fluid bed systems for chemical synthesis

More than 100 fluid bed reactors for chemical synthesis and comparable processes using Group A powders have been installed and operated successfully since the late 1940s, with some reactors having an inside diameter at much larger than 7 m (23 ft). This article discusses developmental milestones leading to their success, the key design differences between chemical synthesis circulating fluid beds and fluid catalytic cracking (FCC), and the author's recipe for successful scale-up. Root causes of typical operational issues of this class of beds are also discussed.

Introduction. Types of gas fluid beds used typically include:

- Fixed fluid bed (FFB), also called bubbling or conventional fluid bed. Examples include phthalic anhydride, acrylonitrile, ethylene dichloride, maleic anhydride and Sasol gas-to-liquid (GTL).
- Dual FFB circulating bed. Examples includes Model IV FCC and others.
- Circulating fast fluid bed (CFFB), using a riser reactor. Examples include Sasol Synthol GTL, DuPont maleic anhydride, TRIG coal gasification, catalytic waste conversion, gas-to-olefin and chemical looping. They are characterized by a riser velocity of < 10 m/sec and very high riser solids holdup.
- Circulating fast bed (CFB), using a riser reactor. Examples include FCC, pyrolysis and coal combustion. They are characterized by a riser velocity of approximately 6 m/sec and low riser solids holdup.
- Moving beds and downers.

Fluid bed chemical synthesis reactions convert gas-phase hydrocarbons to high-value-added products. Reaction is typically limited to around 450°C (840°F). The more common mode of operation is with gas-phase oxygen supplied as air, enriched air or oxygen. When using gas-phase oxygen, operating pressure is usually limited to around 3 barg (45 psig) due to safety concerns and also because gas-phase oxygen reactions are favored by reduced pressure. Higher pressures may be used safely when oxygen is provided in solid form or when there is no oxygen, as occurs in some processes with comparable characteristics, such as the Sasol circulating or bubbling GTL process.

Since the mid-1940s, more than 100 FFB reactors using Group A powders have been installed and operated successfully for chemical synthesis and comparable processes around the world. Their success was based on the following developmental milestones:

- Correct overall range for catalyst size (wide size distribution, high levels of 0 micron–40 micron)
- Figuring out how to perform research at the lab to obtain data for scale-up correctly
- Correct fluidization regime(s) to use
- Proper gas distributor design (many holes distributed evenly with adequate pressure drop)
- Placement of coils inside the reactor and the process and mechanical design parameters of the coils.

Use of CFFB systems using Group A powders for chemical synthesis also dates back to the 1940s, but their use has been less successful than FFBs due to more complex design, higher risk when gaseous oxygen is present, and limited window of operation when internal coils are needed (metal erosion). CFFBs have been used or assessed for several types of large-scale processes:

- GTL by iron oxide reduction (solid-phase oxygen)—Sasol Synthol process
- Hydrocarbons to oxygenated chemicals by adsorbed oxygen (solid-phase oxygen)—DuPont butane to maleic anhydride
- Hydrocarbons to oxygenates with air, enriched air or oxygen (gas-phase oxygen)
- Natural gas to olefins by adsorbed oxygen (solid-phase oxygen)
- Natural gas to olefins with air (gas-phase oxygen)
- Non-slugging gasification of coal with oxygen (gas-phase oxygen)—TRIG process
- Conversion of solid waste to chemicals or syngas
- Chemical looping.

FIG. 1 depicts two CFFB systems presented in literature.[1] The sketch on the left represents the Sasol Synthol process for conversion of syngas to gasoline using iron oxide as catalyst. The iron oxide is reduced as conversion progresses. The sketch on the right is the concept piloted and commercialized by DuPont to convert butane to maleic anhydride using adsorbed oxygen. Both of these systems circulate a solid between two or three vessels, consisting of a riser reactor, a stripper/solid separator and, when required, a regenerator. The regenerator may operate as an FFB or a riser.

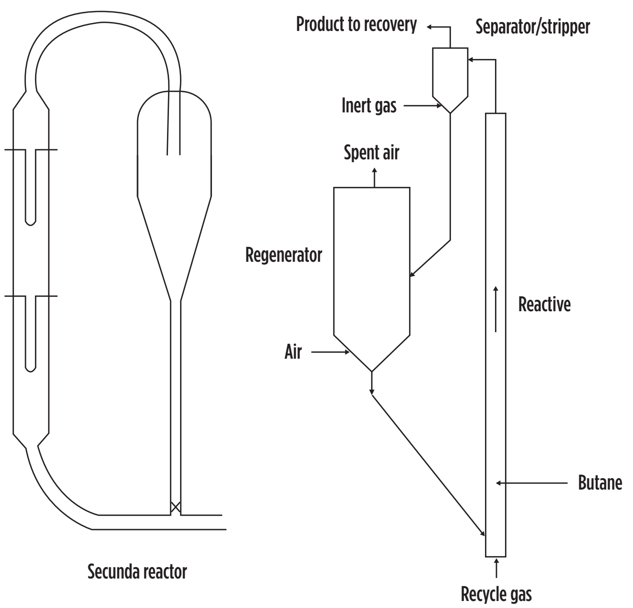

FIG. 1. Examples of CFFBs.

CFFBs have been assessed both in concept and through demo-scale production plants for the conversion of natural gas directly to olefins by most major gas field producers. So far, however, the economics have not justified further development—the main barrier being process yield.

CFFB vs FCC. A CFFB has many design aspects similar to that of an FCC, which is referred to here as a circulating fast bed (CFB). A Group A powder chemical synthesis CFFB will also contain a reactor that operates as a riser, but it should not be designed using FCC design guidelines for the following reasons:

- Use of oxygen gas in the reactor side of the CFFB poses safety considerations with respect to the design of the reactor in the event of loss of solids circulation. This concern is not present in FCC. Deflagration may occur in the CFFB reactor due to a drop in solids holdup caused by loss of solid circulation.
- Chemical synthesis catalyst may cost up to $110/kg ($50/lb) with a life of many years compared to less than $8/kg and a few months of life for FCC catalyst. In CFFB, the unit is typically designed to minimize attrition. In FCC, attrition of the catalyst is not a major design concern.
- The diameter ratio of regenerator to standpipe in modern FCCUs is very large. The ratio in a CFFB with the same size standpipe is significantly smaller as coke production is much less in CFFBs, or not present at all. Most regenerator standpipe entry designs used in FCC do not translate well to a CFFB.
- FCC uses atomized liquid feed in the reactor. A CFFB for chemical synthesis is likely to use gas feed, which creates challenges on how to mix the feed with circulated catalyst in the reactor bottom, especially if the reaction involves the use of oxygen (gas or solid phase).
- The FCC reactor is endothermic. The FCC regenerator is exothermic and may or may not require heat removal. The CFFB may require heat removal in both the reactor and the regenerator. The use of a catalyst cooler commonly used in modern, large-scale FCCUs is not practical in the CFFB reactor side, or often even for the regenerator.
- Fast separation of gas and solid at the exit of the riser in the FCCU is critical to minimizing gas yield. This separation is not as important in most, if not all, CFFBs.
- CFFBs require special considerations on how the reactor exit is connected to cyclones due to the high cost of the catalyst.
- Average density in the FCC riser is about 80 kg/m^3 (5 lb/ft^3). In the CFFB, riser density is typically more than 160 kg/m^3 (10 lb/ft^3). The CFFB reactor also operates at significantly lower velocity than the FCC riser. Moving such a dense mixture vertically up a tall riser at low velocity is not an easy task.
- FCC catalyst pores contain high-boiling-point components that are difficult to strip. CFFB catalysts typically does not have this issue, so stripping in the CFFB is generally easier.

Regenerator temperature control options. Approaches to controlling the temperature of the regenerator vary depending on the type of process being considered, ranging from external catalytic coolers for FCC to internal cooling coils or other approaches for chemical synthesis (**TABLE 1**).

Riser solid holdup. **FIG. 2** may be used to estimate CFB riser solid holdup as a function of velocity and solid flux and applies to solid particle density of approximately 1,100 kg/m^3–1,750 kg/m^3. The vol% of solids can be obtained from the graph and multiplied by particle density. Commercial CFFBs typically operate with solids holdup above 10 vol% and fluxes over 300 $kg/m^2/sec$.

Scale-up. The author's experience suggests that there are no hydrodynamic reasons to limit the capacity of a fluid bed using Group A powder that has proper solid size distribution, operating velocity and design of internals. The factors that limit CFFB unit capacities (reactor size) are typically as follows:

- **Plot space:** The structure space (width, length and especially height) needed to connect three vessels (reactor, stripper, regenerator) together in a CFFB loop becomes excessively large as equipment size is increased.
- **Cost factors:** Shop fabrication is usually cheaper than field fabrication. Vessel diameters are, therefore, limited to about 4.5 m (15 ft) due to over-the-road accessibility with shop fabrication. Since FFBs have a single vessel, the structure height changes only a little with the size of the reactor. As a result, single-train field-fabricated FFB reactors of 7 m (23 ft) and larger are commonly used.

The pilot plant, demo plant or first commercial plant of a new process does not need to be optimal. The plant should work reliably and prove that the process concept is commercially viable. Optimization can be achieved by tweaking the second commercial plant. Admittedly, this approach may require some calculated risk. It is the faster approach to demonstrating a new Group A-based fluid bed process (CFFB, CFB or FFB). The alternative is to resolve all issues, thereby extending "time to market," during which period engineering and research hours are spent increasing the "cost to market."

TABLE 1. Parameters impacting regenerator cooling options

Feeds	Product	% coke	Solids circulation	Regenerator airflow	Regenerator temperature
Raw gasoline	FCC products	Base	Base	Base	~ 700°C
Light hydrocarbons + gas O_2	CxHyOz	<< Base	~ Base	<< Base	< 450°C
Light hydrocarbons + adsorbed O_2	CxHyOz	<< Base	> Base	<< Base	< 450°C
Syngas + solid O_2	Gasoline	0	>> Base	0	–
Natural gas + gas O_2	Olefins	0	>> Base	0	–
Natural gas + adsorbed O_2	Olefins	0	>> Base	<< Base	> 700°C
High-sulfur gasoline	Low-sulfur gasoline	0	<<< Base	<<< Base	< 450°C
Confidential	Confidential	>> Base	<<< Base	<<< Base	~ 600°C

The development of the Z-Sorb low-sulfur gasoline process is an excellent example of how collaboration between research and engineering—and taking some calculated risks—can expedite the time to market of a new process. The author was part of the team that designed, built and successfully demonstrated a 6,000-bpd unit within 16 mos of the project award, allowing ConocoPhillips to license the process before the short window of opportunity disappeared. Engineering started on Day 1, based on limited bench-scale data, and progressed while bench-scale research continued in parallel.

If the research at the bench scale is done correctly, then all the information needed to design a commercial, high-fines Group A reactor can be obtained. The only reasons for using a pilot plant is to build confidence (an admittedly valid reason) and to verify issues that cannot be addressed at the bench scale. Some of these issues may be addressed at much lower cost using cold mock-ups.

CFD capabilities in modeling gas/solids systems have increased tremendously and will continue to do so. CFD is a useful tool for supporting successful scale-up of new, large-scale fluid bed reactors using Group A powders. Unfortunately, some organizations rely on CFD results too heavily, without asking if the results make sense.

Many articles have been written on how solids behave inside a riser and how this may impact reactor yield and gas back-mixing. Many such articles were written years ago as bubbling beds were being introduced. Obviously, the design of the riser exit can cause recirculation of solids, and therefore gas. As for the rest of the riser, the higher the solids concentration of a riser, the more "well-behaved" it becomes. In addition, although chemical plants are designed to operate at full capacity, they sometimes must operate at reduced rates for extended periods of time. If a new process cannot handle turndown using riser technology, then perhaps the use of a riser is not the best choice?

The hydrocarbon processing industry witnesses an occasional paradigm shift that allows novel and untried approaches to become viable. However, it is also true that almost every process used to produce a chemical via the application of fluid beds has already been examined in the past, in some form or another.

Recipe for successful scale-up. A list of lessons learned for the successful commercialization of a new, large-scale FFB, CFB or CFFB process for chemical synthesis includes:

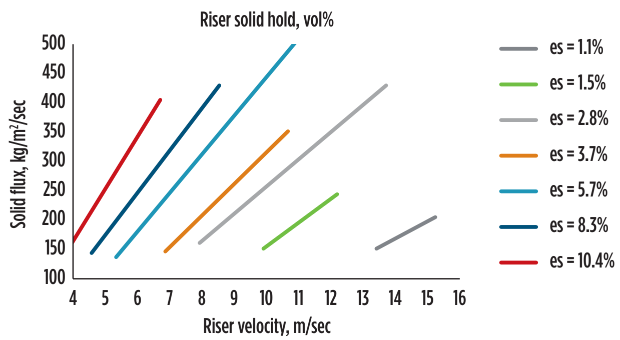

FIG. 2. Riser solid holdup.

- Catalyst with acceptable characteristics is available (this is a given)
- Bench-scale research is carried out correctly; this is easier said than done. Many organizations do not obtain data correctly. The challenges we face are highlighted by the use of partial oxidation reactions represented by Eq. 1:

$$CxHy + O + (N, Cl, etc.) = Product + H_2O + COx \quad (1)$$

Partial oxidation reactions are favored by reduced pressure. The lower the operating pressure and the lower the concentration of hydrocarbon in the total feed, the higher the per-pass conversion and per-pass yield. Testing such a process using a low-pressure or an atmospheric pressure unit, and focusing on adjusting only the concentration of the hydrocarbon, masks this effect. This results in lower-than-expected yields when the unit is scaled up in the pressurized pilot/commercial plant.

Some partial oxidation reactions can also suffer from the formation of "color bodies." These are trace byproduct(s) formed in ppm concentrations or less when the catalyst is not at its optimal condition. Color bodies must be removed, sometimes with great difficulties, to ensure that the end-user product meets the required color specifications (hence "color bodies"). The conditions under which color bodies are formed can be found only through testing at the lab or pilot plant.

FIG. 3 shows the impact of reactor size on yield for a

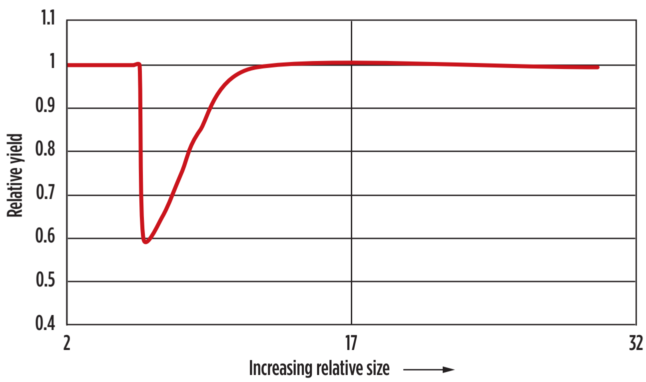

FIG. 3. Impact of scale on process yield.

high-fines Group A powder process. The data is masked, as it is proprietary, and is reported relative to the smallest unit tested. Constant pressure, temperature and gas contact time were maintained. **FIG. 3** shows that yield remains essentially constant as size is increased, except at one scale, due to hydrodynamics caused by bubbles. Not all Group A powder processes show this behavior.

- Address issues immediately, rather than ignoring them or wishing them away.
- Take a total engineering approach to the design of the pilot plant and the first commercial plant. The new reactor always receives the most attention in the design phase, but it cannot operate if the front or back end of the process is designed or built poorly and is incapable of reliable operation.
- Engineering should collaborate with the research team as soon as the viability of the process is validated in the lab, and adequate and repeatable yield data is obtained to start the first-pass concept plant design. This exercise often leads to the discovery of gaps in research that can then be addressed before committing to the pilot plant. These gaps will eventually be discovered and will need to be addressed. Unfortunately, this discovery tends to happen after the engineering firm is hired and staffed up to design the pilot plant. Engineers charging hours will end up working at a less-than-optimal pace while waiting for the missing information, thereby increasing the project cost and schedule.
- Select the correct engineer for the pilot plant design. The local full-service or boutique firm may be a lower-cost option for the design and construction phase, but it may not be the correct choice. Remember, this is fluidization and solids processing. Despite the advancements in the field, it is still somewhat of a "black art." If the technology developer is forced to spend money to fix problems before steady-state pilot testing can begin, then any perceived cost savings during the design and construction phase may be spent multiple times over. In addition, by this time, the process owner will have operating staff on the payroll just waiting around, which becomes expensive. Not every company has deep pockets to ride through such an unfortunate event or reassign idle operators to other activities.
- Correct process and mechanical engineering should be carried out for the pilot and the first commercial plant reactor, other fluidized vessels and their internals.
- Not spending the correct amount of money is another caution. There are reasons why the cost for 2-in. metal-seated full-port ball valves and an actuator range from \$3,000–\$25,000 (U.S. suppliers' data only).

Common operational issues in chemical synthesis reactors. Common issues related to the operation in chemical synthesis reactors include:

- Poor yield/selectivity, caused by:
  - Reaction chemistry not studied correctly/insufficiently
  - Incorrect catalyst size distribution (too coarse/too fine)
  - Incorrect reactor velocity (too low)
  - Poor gas distributor design or damaged distributor
- Reactor instability, caused by:
  - Reactor velocity too high (high velocity can mean low bed density and runaway reactions)
  - Coil layout too restrictive (solid motion must be reasonably free to dissipate the heat generated)
  - Incorrect distribution of coils within reactor (coils must cover an adequate portion of the reactor)
- Catalyst attrition, caused by:
  - Catalyst is too soft
  - Incorrect gas distributor design (jet velocity too high)
  - Incorrect cyclone design (cyclone velocity too high)
  - Incorrect gas distributor fabrication/installation or insufficient quality check (either new or repaired)
- Catalyst deactivation, caused by incorrect grid plate design
- Erosion of internals, caused by incorrect placement of internals relative to gas jets
- Filter operation, caused by:
  - Design solid loading to filter too low
  - Design filter face velocity too high
  - Design filter cake density too high
  - Design gas inlet pipe layout incorrect
  - Blowback gas temperature too low, causing pore plugging.

Recommendations. Commercializing a new Group A fluid bed process requires extensive research to support catalyst development and prove reaction chemistry, but its commercial success ultimately relies on the capabilities of the engineers that scale up and design the first commercial unit for the process. As such, engineering provides a critical and complementary function to research in the development of any new multiphase process, of which fluidization is one of the more important types. HP

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HP | Environment and Safety

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Consequence modeling and risk analysis for safeguarding against jet fires

Due to their high momentum and erosive nature, sustained jet fires are one of the most hazardous types of fires that can threaten the integrity of process facilities. Identifying jet fire risks and providing adequate protection early in the design are crucial for mitigating—in the most economical manner—the consequences of, and the potential for, the escalation of major accident events, while ensuring compliance with applicable laws and regulations. Although guidance and methodology are available for determining protection against pool fires (e.g., API RP 2218), there is no universally recognized code or standard that addresses jet fire risks. A case study involving a propane dehydrogenation (PDH) unit is presented here to demonstrate the value of consequence modeling and risk assessments to economically determine an adequate level of protection against credible jet fire scenarios via a combination of prevention, control and mitigation measures, as early as in the definition phase of a project.

Jet fires. Jet fires pose a serious hazard to hydrocarbon processing facilities, with the potential to escalate into a major incident. A jet fire occurs when a flammable fluid (such as a gaseous, flashing liquid or two-phase and pure liquid inventories) is rapidly released from a pipe or orifice and immediately ignites. The fire's heat intensity poses a hazard to personnel and can cause damage to unprotected equipment due to direct flame impingement. Jet flames also dissipate thermal radiation away from the flame's visible boundaries and transmit heat energy that could be dangerous to life and assets.[1]

In 2007, a jet fire from a pressurized liquefied petroleum gas (LPG) release at the Valero McKee refinery in Sunray, Texas, caused very rapid heating and the failure of unprotected structural steel, resulting in the collapse of a pipe bridge, which greatly increased the magnitude of the fire. The incident resulted in injuries to three employees and a contractor, with direct losses attributed to the fire exceeding $50 MM.[2] The U.S. Chemical Safety and Hazard Investigation Board's investigation of this incident found that a jet flame (approximately 77 ft) impinged on a pipe rack that was located outside of the fire scenario envelope (approximately 50 ft), which considered pool fire risk only.

Current industry practice focuses on analyzing the flame length, thermal radiation intensity extent of impingement and duration of potential jet fires, as well as the need for passive fire protection (PFP), emergency depressurization and other mitigation options.[1] The requirement for projects to drive down costs often does not support the universal application of jet-fire-rated fireproofing, and, thus, warrants some form of risk assessment to create practical solutions. A full quantitative risk assessment (QRA) may yield the optimum recommendation for jet fire protection; however, the level of detail required for a QRA is often not available during the early stages of a project, such as front-end engineering and design (FEED).

Although various industry references[3,4,5,6] provide some form of guidance and background material for consideration, there is no prescriptive method to address jet fire risk. Previous work summarized proposed alternatives for conducting full QRAs and their respective limitations, and provided a methodology suited for offshore fast-track projects.[7] However, the methodology does not address the lack of industry-standard data required to conduct frequency analysis for onshore projects. The goal of this article is to present a methodology to determine appropriate jet fire protection in FEED projects by applying a semi-quantitative risk assessment that involves a combination of consequence modeling and simplified hazard identification or a coarse QRA, where available. The methodology is demonstrated with a case study involving a PDH facility.

Method. A summary of the jet fire risk assessment methodology applied in this case study is shown in **FIG. 1.** The risk assessment is initiated by the available information from the coarse QRA (e.g., high-risk areas and pre-identified failure scenarios). In the absence of a coarse QRA, a process hazard analysis—such as hazard identification (HAZID) or hazard and operability (HAZOP) studies—or a simplified hazard identification technique can be applied where process hazards are identified based on hydrocarbon inventories, operating conditions, and equipment type and location. This information is validated by project data—including process flow diagrams, piping and instrumentation diagrams (P&IDs), and heat and material balance—and then utilized for consequence analysis (e.g., jet fire modeling and impact assessment) to propose the recommended preventive, control and mitigative measures.

The jet fire scenarios were modeled using proprietary software[a], based on the following general parameters:

- **Release size (25 mm):** The release size can range from pinhole to fullbore rupture (**TABLE 1**). Different leak sizes have different implications with relation to impact potential. Guidance is available for selecting generic hole sizes for performing the fire hazard analysis.[3] The small and medium releases have a longer duration, along with a greater potential for escalation and domino effects. Large and rupture releases may have larger hazard zones, but the shorter duration is less likely to cause structural failure. For this case study, the medium release category was selected, since the event is still large enough, and the duration long enough, to cause the greatest impact.[8]
- **Release direction (horizontal):** The accidental release can occur in any direction (horizontal, vertical or angled releases); however, the most hazardous ones are horizontal releases because the resulting jet fire will be closer to nearby equipment, with a high probability of impairment. Per the jet fire correlations implemented in the proprietary software, the flame length and width will decrease with an increase in the angle from the horizontal from 0° to ±90°.
- **Release elevation (1 m from grade):** The proprietary software documentation recommends that the release elevation should be set to greater than zero (e.g., 1 m) for two reasons:
    1. This release elevation is more realistic since most releases are not at ground level.
    2. This elevation provides better modeling of droplet behavior for continuous releases.

  If a continuous release is at ground level, then the program assumes that all liquid rains out immediately to form a pool—whereas, if the release is above ground level, the program models the evaporation and trajectory of the droplets inside the cloud, and the proportion that rains out will be lower and more realistic.
- **Weather conditions (2F and 5D):** The weather condition is generally selected based on the wind rose of the facility or data from the nearest meteorological office weather station. The two most common wind/weather conditions used in the UK—accepted by the UK Health and Safety Executive agency for Control of Major Accident Hazard (COMAH) reports—are:
    - Wind speed at 2 m/sec and a Pasquill-Gifford stability class F, representing very stable weather during nighttime (2F)
    - Wind speed at 5 m/sec and a Pasquill-Gifford stability class D, representing neutral weather (5D).

TABLE 1. Jet fire release categories and hole sizes

Release category	Release size range
Small	0.1 in.–0.4 in. (3 mm–10 mm)
Medium	0.4 in.–2 in. (10 mm–50 mm)
Large	2 in.–6 in. (50 mm–150 mm)
Rupture	Full bore (equipment diameter)

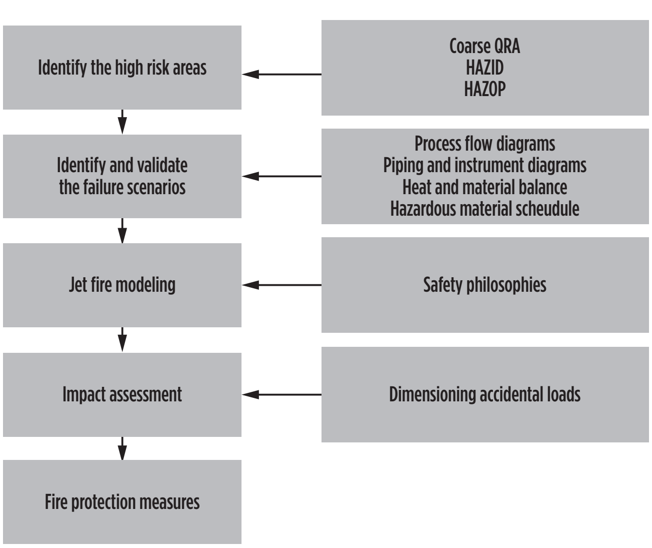

FIG. 1. Proposed jet fire risk assessment methodology.

The impact assessment was carried out by comparing the jet fire modeling results with failure time criteria. As part of a risk-based approach, dimensioning accidental loads (DALs) provided a convenient metric to assess appropriate levels of fire protection. Typical load measures include the duration of flame impingement, exposure to radiation levels above a defined value, heat transfer or temperature increases. Additionally, exposure duration is the simplest and most widely reported DAL, which was used in this methodology.[9] The jet fire dimension at time zero is generally used in the QRA to determine the number of immediate fatalities; however, for asset protection, longer fire durations are used. The failure of the unprotected vessel or load-bearing member/support is assumed to occur within 5 min in case of jet fire impingement.[10]

Protection measures were recommended based on the consequence analysis results (i.e., jet fire modeling and the impact assessment). The protection measures were applied by following the hierarchy of control.

Results and discussion. In this case study, the PDH facility utilizes a licensed technology designed to produce up to 750,000 tpy of propylene from propane feedstock. A PDH flow schematic is shown in **FIG. 2.** During normal operation, the PDH facility contains varying hydrocarbon compositions at pressures as high as 22 barg, with an estimated 1,400 t of total flammable liquid inventory.

A jet fire protection philosophy usually involves three elements: system de-inventory, passive fire protection and active fire protection. However, due to operational constraints inherent to the process, a system de-inventory process to minimize the risk of sustained jet fire impingement on either process equipment or structures, and to also mitigate subsequent escalation, was not practically achievable nor economically viable. Due to the licensed technology employed, the major impediment to the successful emergency de-inventory of the facility was that it could only be executed after a considerable time delay upon initiation of emergency shutdown to bring the facility to a safe state. Further, the inherently large flammable inventory within the process necessitates the application of a

time delay during shutdown to avoid a disproportionately large flare capacity and to prevent possible equipment damage due to the resultant overpressure. It was also recognized early in the design that, because of the directional nature of jet fires and the inherent congestion of modular process units, active fire protection through firewater monitors would not adequately address the jet fire risk. Therefore, jet fire protection for the PDH facility was focused on the application of risk-based prevention, control and mitigation approaches.

To determine which facility assets required protection, consequence modeling was performed to characterize the severity of jet fire scenarios. As the PDH process facility involved an overwhelming number of sources from which the jet fire could originate, it was necessary to determine the minimum number of jet fire models that adequately represented the risk of escalation in the facility. In this case study, pertinent data from the coarse QRA that was developed during the pre-FEED phase of the project was utilized—hence, the semi-quantitative approach. The coarse QRA contained information on which process sections were considered to be the high-risk areas based on the location-specific individual risk, considering various fire and toxic release events. Seven jet fire sources were found to represent the high-risk areas for the facility (**TABLE 2**).

Case study analysis. Once the jet fire scenario models were developed, the jet fire contours were compared against the equipment layout plans and the three-dimensional model of the facility to determine which assets could potentially be impacted. **FIGS. 3** and **4** illustrate the two-dimensional and three-dimensional jet fire contours, respectively, from Source 4 in the product splitter section. For this example, a systematic assessment was carried out, taking into consideration various options for prevention (e.g., equipment design and flange guards), control (e.g., fire and gas detection) and mitigation (e.g., fireproofing and active fire protection) measures. The fi-

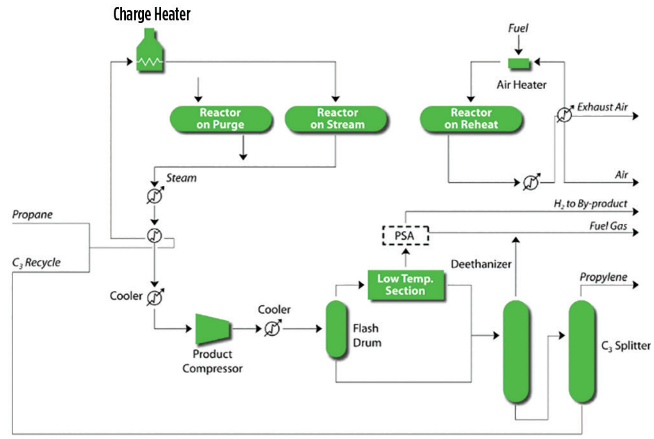

FIG. 2. Flow schematic of a PDH process.[11]

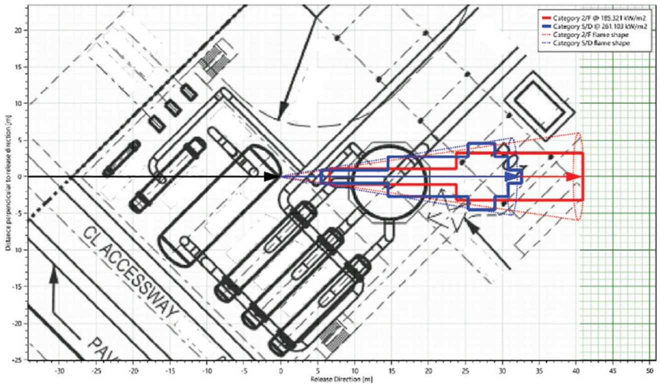

FIG. 3. Two-dimensional view of jet fire contours in the product splitter section.

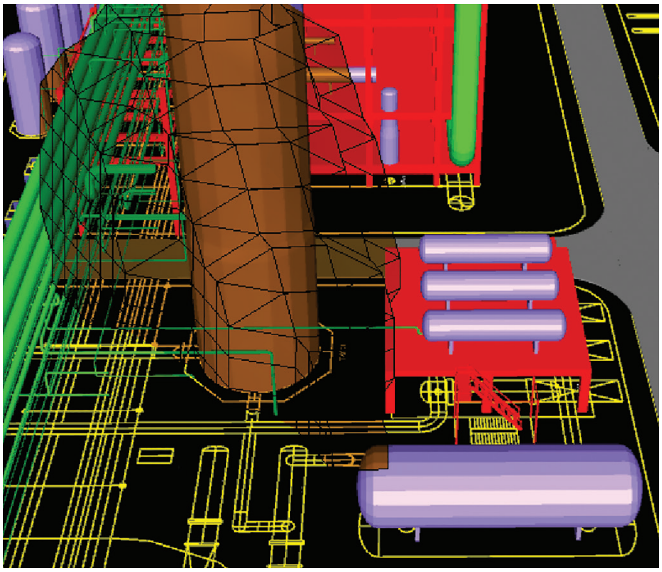

FIG. 4. Three-dimensional view of jet fire contours (category 5D) in the product splitter section.

TABLE 2. Jet fire sources within the PDH facility's high-risk areas

Jet fire source	Process section	Temperature, °C	Pressure, barg	Inventory, kg	Phase
1	Propylene refrigeration	73	17.5	7,800	Vapor
2	Product compression	92	12.1	9,400	Vapor
3	Product compression	16	11.3	9,400	Vapor
4	Product splitter section	49	19.5	141,700	Vapor
5	Product splitter section	59	19.8	227,600	Liquid
6	Deethanizer	-11	5.5	1,100	Vapor
7	Low-temperature section	-20	10.3	1,200	Vapor

nal recommendation was that all process equipment holding a large hydrocarbon inventory within a 40-m radius from Source 4 should be provided with suitable protection through a combination of prevention, control and mitigative measures against jet fire impingement to minimize the risk of escalation. TABLE 3 provides a summary of the jet fire modeling results.

> Identifying jet fire risks and providing adequate protection early in the design are crucial for mitigating the escalation of major accident events, while ensuring compliance with applicable laws and regulations.

Case study—Safeguards and protection measures. Remotely operated shutoff valves (ROSOVs)—also known as emergency isolation valves (EIVs)—are equipped with actuators, and are configured to be quickly and reliably operated from a safe location, such as a well-sited control room. ROSOVs should be used in facilities where fast and effective isolation is needed to reduce the impact of major hazardous releases.[12] In this case study, ROSOVs had been applied in some process sections. However, if the application of ROSOVs is not accompanied by timely depressuring/de-inventory actions, its effectiveness in fully mitigating the risk of incident escalation from jet fires is limited, especially for relatively large, isolated inventories.

An assessment was carried out to determine the provision of ROSOVs by following guidance from the UK Health and Safety Executive.[13] ROSOVs can provide rapid isolation of plant items containing hazardous substances in the event of a failure of the primary containment system—including, but not limited to, leaks from pipework, flanges, valves and pump seals. A ROSOV offers risk reduction benefits by allowing the isolation to be made from a safe distance (such as via a remote push-button system or from a control building) when combined with fire and gas detection systems.

Based on the study, it was determined that, because of large inventory and the physicochemical properties of propane, the pressure could not be sufficiently reduced, even when the inventory was blocked off through ROSOVs—therefore, the application of ROSOVs was determined to be ineffective. For example, jet fire Source 5 is a candidate for the addition of ROSOVs, due to its high operating pressure (19.8 barg) and large flammable liquid inventory (nearly 228 t). However, as Source 5 encompasses a large pressure vessel and large pipework (up to a 46-in. nominal size), the provision of ROSOVs was estimated to reduce the largest isolatable inventory by less than half of the initial volume. This translated to a jet fire release duration of more than 12 hr, which still posed considerable jet fire risk to the facility. The application of ROSOVs was still recommended, although it was concluded that it did not sufficiently mitigate the risk on its own. Consequently, other combinations of risk reduction measures were explored.

Risk of credible jet fires and congestion due to modular design posed a unique challenge from a fire protection perspective. Therefore, a combination of prevention, control and mitigation measures were required to adequately address the risk of jet fire, along with methods for escalation prevention. Taking into consideration the control hierarchy, a multidiscipline assessment was carried out to systematically apply protection measures. Various options were explored to reduce the jet fire frequency and to bring this risk to an acceptable level where no additional protection measures would be required. These measures included the possibility of improving the equipment design through such measures as using seal-less technology for pumps, providing increased thickness for process piping, and reducing the number of flange connections and flange guards.

The fire and gas detection systems were already specified; therefore, it needed to be ensured that the potential leak sources were within the coverage of flame detectors, so that any potential jet fire could be quickly detected and control measures could be initiated. Additionally, close-circuit television coverage was provided, so that any jet fire incident could be rapidly detected and protective measures employed. While some of the equipment were already specified with insulation for process reasons, they also required protection against jet fire impingement.

Active fire protection through a water-spraying deluge system was explored; however, due to its ineffectiveness against jet fires, it was decided to upgrade the process insulation to fireproofing, ensuring that the fireproofing material was suit-

TABLE 3. Jet fire modeling results summary

Jet fire source	Process section	Maximum flame length, m	Maximum frustum tip width, m	Maximum radiation intensity, kW/m²	Full release duration, min
1	Propylene refrigeration	16	2.4	173	210
2	Product compression	12	1.5	156	400
3	Product compression	13	1.8	153	360
4	Product splitter section	41	11.9	261	370
5	Product splitter section	19	2.8	183	> 1,500
6	Deethanizer	10	1.4	108	60
7	Low-temperature section	10	1.1	125	60

able for jet fire protection. Structural steel within the jet fire impact zone was specified to be fireproofed up to 12 m from grade with a jet-fire-resilient fireproofing material. Secondary protection was provided through remote-controlled firewater monitors, ensuring the application of copious amounts of water for cooling purposes for protection against radiation from high-intensity jet fires.

Takeaways. The methodology presented in this case study brought forth the recommendation to apply a combination of protective measures based on the evaluation of coarse QRA data and consequence modeling. A combination of preventive, control and mitigative measures was specified, taking into consideration significantly large hydrocarbon inventories and congestion due to the modular design of the plant's process units. Although the recommendation of jet-fire-rated PFP to multiple assets early in a project (e.g., the FEED phase) is generally not considered cost effective and is not preferred by site operations and maintenance teams, its application in this case study was justified by the following:

- There was a lack of effective de-inventory capabilities within the PDH facility.
- Insulation was already specified in some of the process equipment, thus requiring only an upgrade to jet-fire-rated PFP.
- Equipment structures and pipe racks in the impact zones were already specified for pool-fire-rated PFP, based on the API 2218 fire scenario envelope approach, therefore only requiring fireproofing material selection suitable for jet fire protection.
- Evidence existed to support that water-spraying systems with design densities specified in codes and standards were ineffective against sustained jet fires.
- There was insufficient firewater monitor coverage due to the modular design of the plant's process units.

As this case study illustrates, credible jet fire risk exists even in onshore projects, especially when the project involves a modular design. Prior to the determination of mitigation measures, it is important that emphasis is first placed on the prevention and control of jet fires by selecting an equipment design that could reduce or eliminate loss of containment events and subsequent fires. Examples of recommended design features include:

- Increased thickness of piping/equipment
- Seal-less technology for pumps handling flammable materials
- Flange guards
- Fireproofing insulation for equipment/pipework suitable to provide protection against jet fires.

Mitigation against jet fires can be achieved by providing the following:

- Jet-fire-resilient fireproofing for equipment structures and pipe racks within the jet fire impact zone
- Adequate firewater monitor coverage, either through grade firewater monitors or elevated remote-controlled firewater monitors
- Water spray with higher discharge densities, where fireproofing for equipment is not preferred due to concerns related to corrosion under insulation. HP

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NOTES

[a] DNV GL's Phast 8.21 process hazard analysis software

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Safety and environmental benefits of reliability engineering

When you boil down the mission of reliability to its bare essence, the job is to deliver maximum operational availability for the least amount of money over the lifecycle of the asset. For this reason, reliability improvement practices often focus on the financial benefits of maintenance, which includes maximizing availability, minimizing cost or reducing failure rates. Meanwhile, many safety programs focus on ensuring that workers follow good practices, such as watching where they walk, being aware of the lines of fire, following personal protective equipment (PPE) guidelines and avoiding pinch points. This creates a business silo mentality, where hazard and operability (HAZOP) studies and failure modes and effects analysis (FMEA) facilitators utilize the same teams to evaluate parallel risks onsite.

Reliability improvement may be the most underutilized and undervalued health, safety and environmental (HSE) mitigation tool that operating units have at their disposal. The same programs that result in fewer unexpected equipment failures also act to effectively reduce HSE risks onsite.

Since these processes are often financially driven, they may represent a more efficient means of managing HSE risks than more traditional safety-driven approaches. Reducing equipment failures, parasitic frictional losses and fugitive emissions inherently increases the profitability of the organization, while reducing environmental impacts and safety risks.

Environmental benefits of reliability engineering. Traditionally, the lifecycle of plant assets has been viewed as a linear relationship, including the designing, producing/configuring, distributing, installing, operating, maintaining and, ultimately, disposing of and replacing the asset (**FIG. 1**). The objective is to optimize investments in upfront costs relative to operational, maintenance, disposal and replacement costs over this lifecycle.

The environmentally minded reliability engineer must rethink lifecycle management to reflect a circular economy that factors environmental impacts into the lifecycle optimization equation (**FIG. 2**). In particular, the environmentally minded reliability engineer must carefully consider lifecycle energy consumption requirements and associated carbon-dioxide-equivalent (CO_2-e) and other greenhouse gas (GHG) emissions, as well as the risks of environmentally hazardous fugitive emissions, asset life and life extension opportunities, and ultimate reusability or recyclability of the asset and the materials from which it was constructed. For most reliability engineers, this comes down to managing parasitic energy losses associated with mechanical and electrical friction and minimizing fugitive emissions. Doing so creates a win-win scenario because parasitic energy losses manifest in the form of heat and vibration. Reducing heat and vibration contributes significantly to the goal of extending the useful life of assets and their components. For example, reducing overall vibration from 8 mm/sec to 4 mm/sec can increase the useful life of a rolling element bearing by a factor of eight times!

A great starting point for the reliability engineer who is interested in reducing environmental impacts is to focus on reducing parasitic frictional losses and fugitive emissions—topics that will be explored here in more detail.

Managing parasitic frictional losses. According to the U.S. Department of Energy, manufacturing, processing and

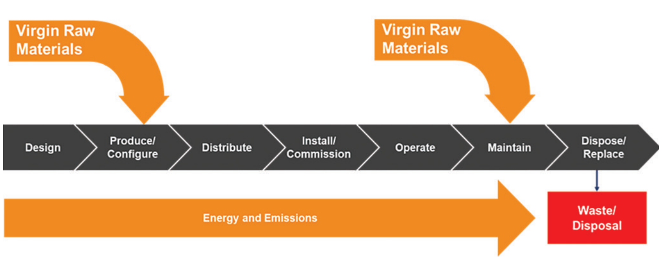

FIG. 1. Conventional asset lifecycle as a linear process.

FIG. 2. Asset lifecycle management reconfigured to reflect an environmentally focused circular approach.

mining operations across a wide array of industries could reasonably reduce energy consumption by 20%–25% by implementing state-of-the-art best practices. To put the opportunity in perspective, industrial plants consume about 30% of the total energy budget in the U.S., and reclaiming this wasted energy could reduce the country's national energy consumption and associated GHG emissions by 6%–7.5%. This article will include a non-exhaustive checklist of some reliability engineering best practices to reclaim some of this lost energy.

The following are ways to employ proactive mechanical best practices:

- Minimize precision mechanical fastening practices
- Follow precision mechanical balancing practices according to ISO 21940 G1.0 for critical assets and G2.5 for the balancing of plant assets
- Precisely align mechanical drive systems:
  - Ensure shaft alignment to 0.3 mm/in. of angularity and 1 mm of offset for 3,600-rpm applications
  - Laser align pulleys, sheaves and sprockets
  - Precisely align conveyor belts
  - Minimize pipe strain
  - Always consider thermal growth in alignment calculations
- Minimize V-belt slip to less than 2%, and use a strobe light to determine slip percentage
- Select the correct lubricant type and quantity:
  - Determine sufficient viscosity and lubricity to minimize boundary contact friction
  - Avoid excessive viscosity, which can produce fluid friction and churning losses
  - Consider the operating temperature and the temperature range, and select lubricants with an appropriate viscosity index
- Keep the in-service lubricant healthy and free of particles, water, air and other contaminants
- Properly size pumps based upon the application
- Operate pumps at peak curve efficiency
- Ensure a sufficient suction-side head to avoid cavitation
- Properly size piping/hosing (diameter and length) and minimize bottlenecks for all fluid conveyance systems, with minimal turns and curves, to minimize fluid friction and churning losses.

The following are ways to employ proactive electrical best practices:

- Use high-efficiency motors—IEC IE4 class motors are about 3% more energy efficient than IE1 class motors (e.g., for a 75-kW electric motor, an IE4 can reduce lifetime energy costs by more than $10,000 vs. an IE1 motor in the same application, as shown in **TABLE 1;** applied to a fleet of 1,000 motors on a typical site, this can result in a staggering $10 MM lifecycle benefit)
- Deploy variable frequency/speed drives where appropriate
- Minimize phase-to-phase electrical unbalance (such as voltage and current), and keep the voltage imbalance to less than 2% (less is best)
- Properly size, install and maintain electrical circuits to reduce heat
- Ensure proper electrical connections in all applications
- Minimize harmonic distortion to less than 3% for any single harmonic and less than 5% total harmonic distortion for systems under 69 kilovolts (kV)
- Minimize reactive power, and make power corrections through synchronous motor control.

TABLE 1. Energy and environmental impact comparison of an IE4 motor vs. an IE1 motor

	IE1, 2 pole	IE4, 2 pole	Differential
Motor kW rating	75	75	0
Assumed load, %	80	80	0
Operating, hr/yr	8,000	8,000	0
Efficiency, %	92.7	95.8	3.1
kWh/yr	517,799	501,044	–16,756
Energy cost/yr at 0.06/kWh	$31,068	$30,063	–$1,005
CO_2 emissions/yr, kg	366,084	354,238	–11,846
Social cost of carbon/yr	$18,304	$17,712	–$592

TABLE 2. Injury rate reduction through precision maintenance

Details	Original	Improved	Differential
Mean time between repairs, mos	12	96	+700%
Corrective maintenance task, FTE	4	4	–
Corrective maintenance task, hr	8	8	–
Corrective maintenance task, worker hr/yr	32	4	–88%
Estimated annual injuries, 4.2/100 FTE	0.0014	0.0002	–86%
Estimated annual cost, $100/hr	3,200	1,000	–69%

Managing fugitive emissions. Fugitive emissions waste energy, and these emissions can cause the unintended release of polluting and hazardous effluent. Compressed air systems represent a great opportunity to reduce energy consumption. Approximately 10% of a typical energy budget is spent compressing air. It is not uncommon for compressed air systems to suffer a 20%–30% leakage rate. This means that a site can leak as much as 2%–3% of its total energy expenditure into the atmosphere. Leaking in other compressed/pressurized fluid systems (such as natural gas; water; and lubrication, hydraulic and steam systems) wastes a great deal of energy. In addition, in the hydrocarbon-intensive industries, flare-minimization strategies have been shown to dramatically reduce waste of methane and other hydrocarbons, and to reduce sulfur dioxide emissions.

In addition to energy conservation, some fugitive emissions have a direct and adverse effect on the environment and on people. For example, during its first 20 yr in the atmosphere, methane is about 80 times more powerful than CO_2 as a GHG before settling down to 25 times more powerful after 100 yr in the atmosphere. Liquid and solid effluents can contaminate watersheds and adversely affect the quality of drinking water. Studies have shown that people living near refineries and other petrochemical facilities suffer rates of cancer that are much higher than normal rates.

Safety benefits of reliability engineering. Many reliability engineers have little to no interaction with the well-regulated

and documented safety processes onsite, such as HAZOP reviews. The reasons may vary, but it often comes down to a simple misunderstanding by site leadership, which often believes that reliability is meant to decrease maintenance costs and increase availability, not address safety concerns. Regardless of where this misconception comes from, the clear and striking distinction between reliability and safety only works to lessen the impact of either initiative.

A reduction in unplanned failure events, along with the systematic processes used to eliminate root causes of failure, can also help to track and improve other metrics related to equipment, including personnel safety risks. Many reliability processes can be applied directly to identify, analyze and mitigate safety risks.

For example, the following case study examines a distillation column system (**FIG. 3**). As with most static equipment systems, the distillation column system has several sources of risk, including:

- Financial risks represented by the production losses associated with system downtimes and maintenance costs to repair the functionally failed equipment
- Safety risks related to processing hazardous and flammable fluids; fugitive emissions can lead to potential environmental reporting

Most operating areas are proficient at ensuring that regulatory inspections are followed for pressure vessels, thereby reducing the risk of atmospheric leaks, fires and explosions. HAZOP studies and other processes may even identify the need for design changes in equipment to meet evolving HSE standards, such as a tandem seal design in the overhead and bottoms pumps, as shown in **FIG. 3.**

With these existing processes in place, it seems like there is little room for reliability tools (such as root cause analysis, FMEA and precision maintenance practices) to have a significant impact. When applied correctly, however, the targeted

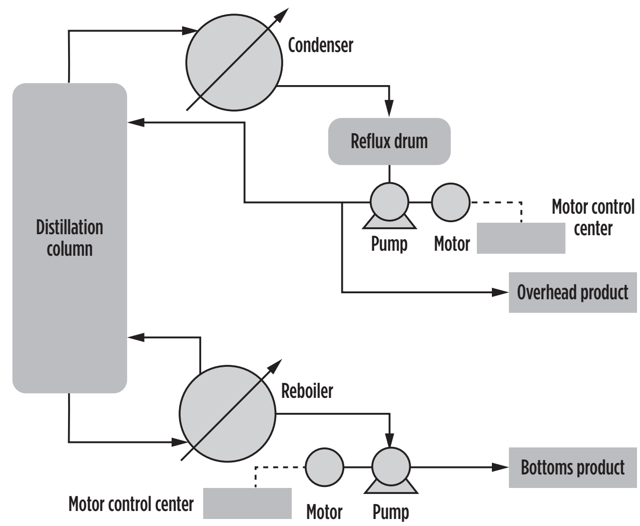

FIG. 3. Distillation column system.

analysis and mitigation steps commonly used in reliability improvement processes are effective at mitigating these risks beyond the scope of traditional safety programs.

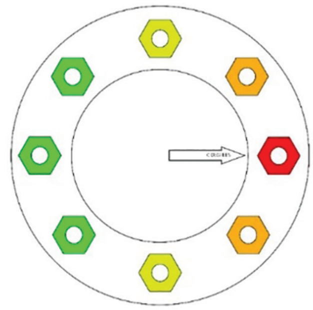

FIG. 4. Improper torquing and tension on a flange.

Static equipment: Distillation columns, reboilers and condensers. Decreasing failures not covered in regulatory inspections reduces all safety concerns related to these static vessels. For insulated vessels, insulation inspections and repairs also decrease losses throughout the system, consequently requiring less thermal energy to maintain optimal process conditions and reducing stress on supporting systems. This can have an impact on steam, nitrogen, air and electrical utilities. The National Insulation Institute estimates that up to 43 MM metric t of CO_2 emissions could be reduced through insulation replacements and upgrades, equating to nearly $4.8 B in energy savings.

Precision maintenance practices, such as detailed and proper fastener selection and installation, also have a major impact on the safety and reliability of these assets. Improperly installed flanges have a significantly increased likelihood of experiencing premature failures. For example, the flange that was improperly torqued (starting at 180°) would likely experience a gasket failure and leak to the atmosphere at 0° (**FIG. 4**). When these failures happen, personnel safety is put at risk twice—when the leak is present and when maintenance technicians repair it.

When high-priority or emergency jobs are executed, proper planning and scheduling are often foregone. This leads to a higher personnel safety risk due to fewer planned parts, materials and tools. Even the most common type of industrial injuries—slips, trips and falls—are at an increased risk when maintenance technicians are traveling to and from job sites more frequently.

The U.S. Bureau of Labor Statistics estimates a non-fatal injury rate of 4.2/100 full-time equivalent (FTE) per year (about 1 injury per 23,000 hr worked). Replacing poor installation and break-in repair with a single precision task could effectively reduce this to less than 3/100 FTE by simply reducing field callouts (**TABLE 2**).

Rotating equipment: Pumps, motors and fans. Rotating equipment (such as pumps, motors and cooling fans) in a distillation system represent different and unique safety risks. Less regulated than pressure vessels, the maintenance and inspection activities on this equipment are most typically determined by the owner's engineer. This often leads to subpar equipment conditions and maintenance, where failures occur well below the expected life of the equipment.

A coupling failure due to high vibration on the pump motor can result in additional safety risks for all personnel in the area. While vibration is inherent to these systems, excessive vibration is most often a result of fasteners, lubrication, alignment or balancing. These conditions that result in the coupling failure can be mitigated proactively through proper installation and maintenance practices and identified early with proper condition monitoring methods.

Regarding bearing life extension, proper precision practices can extend useful life by up to eight times. Fastener failures can result in a sudden and violent dissipation of mechanical energy. In addition to immediate safety risks to personnel, this can result in severe cascade failures of the system, potential releases to the atmosphere and environmental hazards. For example, in Norway in December 2020, a failed steam shutdown valve caused a turbine to overspeed, leading to a catastrophic failure and fire. Failures to identify, classify and mitigate critical equipment led to a lack of proper maintenance procedures and practices. Luckily, only severe financial losses were incurred on this operating unit because of this catastrophic failure.

Electrical equipment: Motors and motor control centers. Hazards often apply to the electrical equipment that supplies the power to rotating machinery. In this case, electrical energy represents the largest safety risk to personnel. Electrical events (e.g., arc flashes) often have the potential to result in severe injuries or fatalities.

Arc flash hazards, which are most dangerous at voltages above 1,000 V, are often not covered under traditional relay schemes. These events commonly occur when maintenance technicians or operators are in the area operating the motor circuit breaker and are switching circuits for normal operation. A single failure in these motor control centers can lead to cascade shutdowns and production stoppages, as the arc flash energy is often high enough to bring adjacent equipment offline. Preventing these events through routine cleaning and testing not only improves worker safety, but also decreases lifecycle costs by preventing catastrophic failures.

Routine cleanings, thermography and ultrasound monitoring in motor control centers can prevent and identify the conditions for fault events before personnel are put at risk. *Industrial Safety and Hygiene News* estimated as many as 30,000 arc flash events per year, resulting in 7,000 injuries and 400 fatalities. In approximate terms, improving electrical testing, maintenance and reliability practices by a minimal 5% across the industry would prevent the following:

- 1,500 failures
- 300 plant shutdowns (up to 3,600 hr of downtime)
- 350 non-fatal injuries
- 20 fatalities.

If safety and shutdown systems are tested on an unplanned basis, poorly maintained equipment can experience electrical failures that can result in immediate dangers to plant personnel. HP

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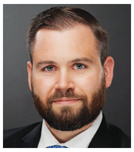

COLEMANN O'MALLEY is a Manager with T.A. Cook. He is a reliability professional, with extensive experience in asset management, risk assessment and implementation of reliability projects in the oil and gas industry. As an electrical engineer with several years of field experience in the refining and chemicals industries, Mr. O'Malley is piloting the implementation and development of digital solutions in asset management.

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Stack gas scrubbing to meet IMO's 0.5% sulfur bunkering requirement

The International Maritime Organization implemented the IMO 2020 regulations on January 1, 2020. As they pertain to sulfur emissions, the regulations require ships without exhaust gas scrubbers to burn either 0.5-wt%-sulfur fuel (very low sulfur fuel oil, or VLSFO) or 0.1-wt%-sulfur marine gasoil (MGO). Ships outfitted with scrubbers can continue burning 3.5-wt%-sulfur heavy fuel (high-sulfur fuel oil, or HSFO). The economics of installing scrubbers depends very much on the price differential between VLSFO and HSFO.

One analysis[1] reported a \$100/t differential as the tipping point above which installing scrubbers becomes cost-effective, although the figure could be as low as \$75/t. Of course, a decision will depend on the specifics of the ship in question with regard to the ease or difficulty of fitting scrubbers and associated equipment, and therefore the capital cost component. As the demand for VLSFO increases, the price differential is likely to rise over the long term (the effects of COVID-19 notwithstanding), not only from the rising value of VLSFO but also from the falling value of HSFO. This will likely make the scrubber option increasingly attractive.

Concern has been expressed by some port authorities about the discharge of sulfur dioxide- (SO_2-) laden seawater from open-loop (once-through) systems when ships are close to ports. The alternative is a closed-loop system in which the solvent contains caustic soda or another alkaline medium to absorb and hold the captured SO_2. Closed-loop systems are more expensive to operate. However, it is fairly straightforward to switch between closed- and open-loop operation, so there is considerable interest in both systems. In addition to monitoring SO_2 from the ship's stack, there are compliance requirements for nitrogen oxides (NO_X); this means that monitoring exhaust gas from ships is becoming more akin to monitoring emissions from refineries.

Some interesting learnings can be gained from the simulation of packed absorbers for this service. One is the huge difference between scrubbing with seawater only vs. seawater spiked with caustic soda. Another is the effect of caustic concentration on performance. Yet another is that scrubbing efficiency is mostly a function of the area of the interface between the solvent and the exhaust gas within the scrubber. A fourth is the negative effect of CO_2 co-absorption, which demands system optimization—maximize the SO_2 absorption and minimize the co-absorption of CO_2. These aspects of shipboard exhaust gas treating are discussed in a later case study.

General considerations. Gas-liquid contacting for the purpose of removing one or more constituents in the gas by absorption into a solvent is usually carried out in a column containing trays or some form of packing (either structured or random). Due to rocking motion (roll, pitch and yaw), treating aboard ships presents special challenges. On land, every effort is made to ensure that columns are closely vertical—a situation that cannot be attained shipboard. Column motion disqualifies trays from serious consideration as contacting devices. Between random and structured packing, structured packing is usually preferred in floating applications (such as FLNG and FPSO) because lateral liquid and gas flows are discouraged by the fact that the structured packing consists of vertically aligned corrugated sheets stacked together. The sheets are an obstacle to the lateral passage of liquid. This helps ensure that gas and liquid flows stay fairly evenly distributed across the column cross-section. In this article, attention is focused on the use of structured packing. However, that is not to say that structured packing is the only contacting device possible. Spray columns are an alternative, but high-performance liquid distributors are still necessary.

A set of spray nozzles uniformly placed over a cross-section near the top of an otherwise empty column can effectively produce a spray of small droplets with large total interfacial surface

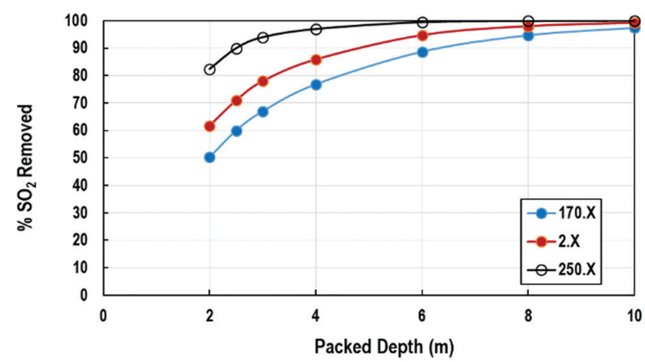
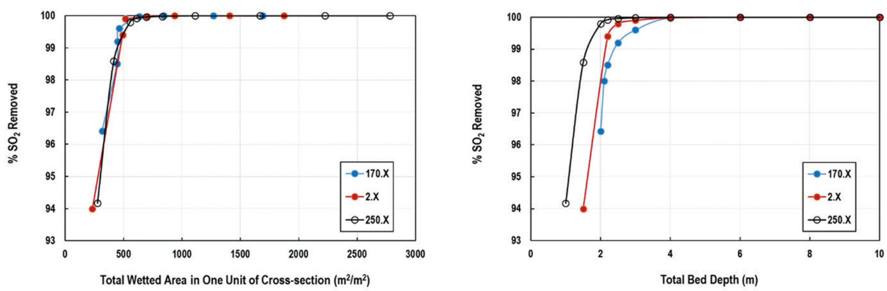
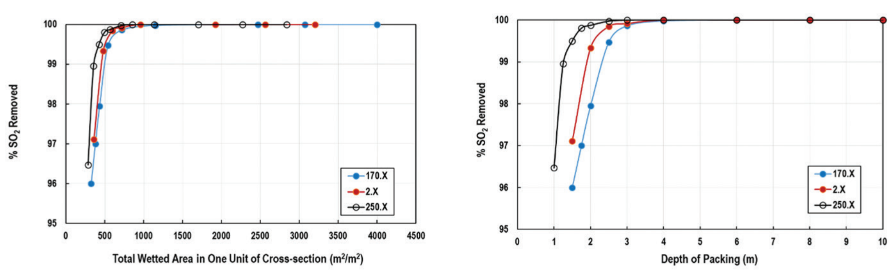


FIG. 1. Dependence of SO_2 removal on bed depth and packing size—once-through seawater.



(a) SO_2 removal depends on total wetted area (b) Dependence of SO_2 removal on packed depth

FIG. 2. Dependence of SO_2 removal on bed depth and packing size—1 wt% NaOH.



(a) SO_2 removal depends on total wetted area (b) Dependence of SO_2 removal on packed depth

FIG. 3. Dependence of SO_2 removal on bed depth and packing size—15 wt% NaOH.

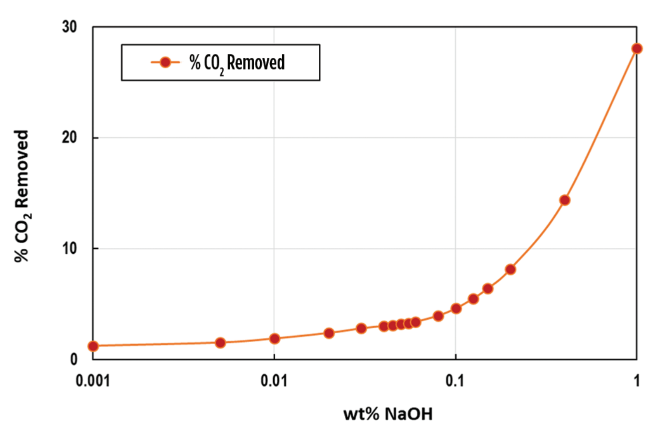


FIG. 4. Using NaOH for SO_2 removal also removes much CO_2. Calculations are for a 1.5-m deep bed of 2.X packing.

area for contacting the upward-flowing gas. However, droplets tend to coalesce with each other and with the walls of the equipment, so it may be necessary to place spray nozzle batteries at more than one level in the column. The other concern with spray contacting is significant carryover of liquid into the emerging cleaned gas. This may not be a big issue if the solvent is seawater; however, the presence of caustic soda is problematic. Carryover can be controlled with mist pads; these would be mandatory using seawater spiked with caustic. As will become apparent from the case study, only a very short packed bed (or, alternately, a short spray section) is needed when caustic soda is used to enhance SO_2 absorption and the upper part of the column can be used to virtually eliminate mist when treating with seawater.

Case study. The specific case is a 76,000-metric-t ship driven by four boiler engines with a combined rating of 67.2 MW (90,000 hp) running on fuel oil, plus two gas turbines rated at 50 MW total. These data correspond to a Queen Mary-type cruise liner. Only the boiler engines are considered in this example.

Two cases are considered: (1) once-through flow of seawater, and (2) recirculating flow of seawater containing some level of caustic soda. To be specific, the absorber column is assumed to contain one of three sizes of proprietary structured packing[a] (M170.X, M2.X and M250.X). The X-series[a] has lower pressure drop than the Y-series[a] because of its 60° crimp angle, which imposes less engine backpressure. The absorber is sized for 85% of flood, and the packed depth is calculated to achieve the same discharge SO_2 concentration as would be in the stack exhaust if VLSFO (0.5% sulfur) was burned. In other words, about 85% of the SO_2 is to be removed by the scrubbing system when burning HSFO. Note that when using caustic soda, much lower sulfur emissions (> 95% removal) can be easily achieved.

A proprietary simulator[b] contains the components needed to fully describe the composition of seawater. TABLE 1 shows the typical compositions of seawater (100°F) and hot engine exhaust gas (800°F) entering the system. The exhaust gas flowrate was slightly over 42,360 standard cubic feet per minute (SCFM), and the absorber was always sized for 75% of vapor flood. Seawater flow was maintained at 2,000 m^3/hr, close to the value needed to remove 85% of the SO_2 in a reasonable packing depth with just seawater. NaOH is listed separately in TABLE 1 because it is added to seawater as though it were a molecular entity. Seawater already has an inherent Na^+ concentration tied up with Cl^- and other ions as part of the charge-neutral ionic soup comprising seawater. Of course, in the mass transfer model, the solvent is treated as an aqueous ionic soup.

SO_2 removal using once-through seawater. Simulations were run over the range 2 m–10 m of total packing depth. FIG. 1 shows how the fraction of SO_2 removed depends on both packed bed depth and structured packing size. Note that the packing size designation is roughly the specific surface area[c] of the dry packing in the units m^2/m^3, and the designation "X" corresponds to a 60° crimp angle, providing the minimum pressure drop. About 85% of the SO_2 is removed in a 2-m bed of 250.X packing. The M2X size (about 200 m^2/m^3) needs a 4-m depth, and 170.X needs between 5 m and 6 m, which means that three times the amount of 170.X packing is needed to achieve the same SO_2 removal as 250.X packing. Column diameter under these conditions is typically 13 ft–15 ft with 170.X, 2.X and 250.X packing, with the finest packing needing the largest diameter.

SO_2 almost always exhibits a very small (usually negligible) equilibrium backpressure, except when the water starts to become nearly SO_2 saturated. Adding caustic soda to seawater not only increases solvent capacity for SO_2, but it also enhances the absorption *rate* itself through acceleration by chemical reaction with the high hydroxyl ion concentration of a very alkaline sol-

vent. If the unspent NaOH concentration is high enough, then solvent pH will remain quite high throughout the column. A potential (and as it turns out, quite real) downside is the fact of significant CO_2 co-absorption. In the engine exhaust, CO_2 is present at approximately 65 times the concentration of SO_2, so despite its lower physical solubility, CO_2 may absorb at a significant rate and parasitically consume a large share of the caustic soda. The higher the caustic concentration, the larger the amount of co-absorbed CO_2 and the larger the fraction of caustic used up by CO_2. It is, therefore, important to use as low of a caustic concentration as possible.

SO_2 removal using seawater spiked with NaOH. Simulations were again run over a range of 1 m–10 m of total packed bed depth using 1 wt% and 15 wt% NaOH in seawater as solvent. **FIG. 2** and **FIG. 3** show the dependence of SO_2 removal on the depth of three beds of 170.X, 2.X and 250.X packing using 1 wt% and 15 wt% NaOH, respectively. It is apparent that SO_2 removal can be described in terms of total wetted area per unit of tower cross-section, regardless of packing size. It is also apparent that SO_2 removal is, at best, a weak function of packed-bed depth itself. Removing exactly 85% of the SO_2 in the engine exhaust can be a challenge using caustic soda because of the extreme sensitivity of removal to bed depth (and wetted area) when substantially less than complete removal is desired. Exceeding SO_2 removal requirements is, of course, a waste of caustic soda; however, in the interest of meeting the removal goal, a ship's crew inexperienced in operating a small chemical process plant can be assured of compliance by exceeding the *required* removal.

Perhaps the more costly consequence is excessive CO_2 removal when no such removal is required. As **FIG. 4** shows, CO_2 removal rises rapidly with the caustic level in the solvent. A 15 wt% NaOH strength will remove nearly 50% of the CO_2 in the stack gas, and even just 1% strength will remove nearly 30% of the CO_2. Given an exhaust of 14% CO_2 and only 0.216% SO_2, between 20 and 30 times more NaOH will be used in removing CO_2 than in recovering SO_2.

Optimizing the system. When seawater is used alone as the solvent, determining the flowrate to achieve a given removal of SO_2 is a fairly simple matter using simulation.[b] The key parameter is the total wetted area in the contacting device. A chart such as that shown in **FIG. 1** is very easy to generate. Although wetted area is a function of liquid rate, under most commercial conditions wetted area and the physical, or dry, area of the packing are similar enough for practical purposes to allow the dry packing area to be used as a reasonable approximation. A large-size packing is preferred because its lower pressure drop results in the least backpressure on the engine. Once the packing has been selected, the needed bed depth can be readily calculated.

The situation is somewhat complicated when the seawater is spiked with caustic soda. Several factors are at play that make caustic addition attractive, but also more complex:

- NaOH provides high concentrations of hydroxyl ion that react with SO_2 and greatly increase its absorption rate (the pH of seawater is typically 8.1, so the hydroxyl ion concentration is only about 2 ppbw). Therefore, with NaOH, much less packing area (i.e., a much shorter column) is needed to remove most of the SO_2.
- CO_2 is also absorbed by caustic soda, and since there is roughly 65 times more CO_2 than SO_2, CO_2 absorption will be a much heavier consumer of caustic unless the packed depth is minimized.
- Keeping the caustic concentration as low as possible will minimize CO_2 co-absorption, but there is a minimum concentration below which the effectiveness of caustic to remove SO_2 falls off rapidly (**FIG. 5**).
- There is nothing to be gained by having the caustic level any higher than necessary. High NaOH just makes the solution more corrosive, more hazardous to handle, and

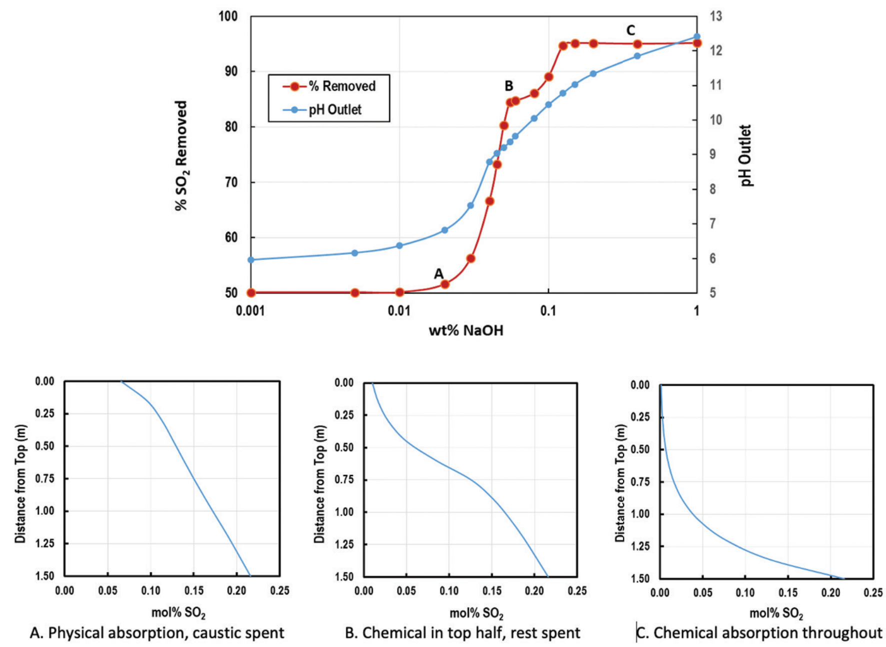


FIG. 5. Types of absorption at various NaOH concentrations. Calculations are for a 1.5-m deep bed of 2.X packing.

TABLE 1. Solvent and gas composition

Seawater, ppmw		Exhaust gas, vol%	
CO_2	90	H_2O	11.08
Na^+	10,800	CO_2	14.052
$SO_4^=$	2,710	N_2	74.478
Mg^{++}	1,280	O_2	0.141
Ca^{++}	410	SO_2	0.216
K^+	399	CH_4	0.00164
Br^+	67	CO	0.0046
Sr^{++}	80	NO_2	0.0256
F^-	1.3		
Cl^-	19,350		
NaOH	Various		

more wasteful when the time comes to discharge the solvent overboard after leaving port.

- The best technical approach is to add caustic to the solution gradually, as it is consumed. This means measuring and adjusting the pH. The scrubber system is a small chemical plant that must be operated at sea by operators relatively unfamiliar with process plant operations. Using a chemically reactive solvent certainly does not make the operation any easier to manage.

In the example given here, the SO_2 removal system can be operated satisfactorily with unreacted NaOH concentrations in the column feed above approximately 0.06 wt%, at which concentration only 2%–3% of the CO_2 is co-absorbed. However, the bed depth is only 1.5 m, whereas the depth needed when using seawater alone is 6 m–8 m for around 90% SO_2 removal. Rather than two columns in parallel, it would make greater economic and physical sense to have a single bed with two feed points, one at the top for seawater only and one at the 1.5-m level (in this case) when using caustic soda. The short bed would be on the bottom for use with caustic, and a deeper bed would be on top, with both beds being operated together when using seawater. Liquid distribution should be via spray nozzles, rather than troughs, to alleviate the effect of sloshing from vessel pitch and yaw. Such a system is not difficult to fully automate with push-button switching from seawater to caustic and with caustic addition under pH control to maintain the effluent solvent above a pH of around 6.0.

The simulator[b] is strictly mass transfer-rate based, and is therefore capable of accurately assessing the effect of the particular packing type, its size and material, and even the brand. The results produced are quantitatively accurate, and the effects of changing parameters on performance can be relied upon. The mass transfer rate-based simulation helps optimize SO_2 seawater scrubber design for sound, environmentally responsible operation. HP

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[1] Miller, G., "Coronavirus is decimating IMO 2020 ship-scrubber savings," *American Shipper,* March 10, 2020, online: https://www.freightwaves.com/news/coronavirus-is-decimating-imo-2020-ship-scrubber-savings

NOTES

[a] Mellapak
[b] ProTreat
[c] Specific area is the surface area per unit volume of packing (area/volume)

PEDRO OTT joined Optimized Gas Treating Inc. in 2018 as a Senior Applications Engineer in Business Development, providing marketing, licensing and technical support to ProTreat and SulphurPro customers. With over 25 yr of experience, he is an expert in acid gas treating, sulfur recovery, gasoline and distillates hydroprocessing, sour water stripping and LNG. He holds a BS degree in chemical engineering from Simón Bolívar University in Venezuela and a specialization in petroleum refining and gas from IFP School in France. In addition to English, he is fluent in Spanish and French.

RALPH WEILAND is Chairman of Optimized Gas Treating. He formed Optimized Gas Treating more than 25 yr ago, and with Australian colleagues developed the ProTreat mass transfer rate-based gas treating simulator, and later the sulfur plant simulator SulphurPro. He holds BASc and MASc degrees, as well as a PhD, in chemical engineering from the University of Toronto.

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Assessment of independent protection layers in an LOPA study—Part 2

Industrial facilities, especially those operating in the chemical, oil and gas and petroleum industries, contain inherent risks in operations due to the processing of materials that are hazardous in nature. Hazards, operability issues, associated risks and their consequences must be accurately identified and analyzed to ensure safe operations. Safety instrumented systems (SISs) are deployed to reduce risk to tolerable or acceptable levels to achieve safe operations.[1] The reliability of safety functions implemented in an SIS are determined by the magnitude of risk reduction required and is expressed in terms of safety integrity level (SIL).

A layer of protection analysis (LOPA) is one of the methods used to determine the SIL of the safety instrumented function (SIF). During the LOPA study, the design is thoroughly examined for one or more independent protection layers (IPLs) in the design to assess whether the required risk reduction has been achieved. The success of the LOPA study depends on proper assessment of the protection layers and their contribution to risk reduction.

In Part 1 of this article (November 2021), three attributes of a LOPA study were discussed: independence, functionality and integrity. The contribution of the remaining three attributes—reliability, access security and auditability—will be discussed here.

Reliability. This core attribute defines the reliability of an IPL, which means that the IPL performs its intended function completely without fail when needed. For an IPL to be reliable, it must be rigorously designed, operated, maintained and tested as per the guidelines/procedures mentioned in the safety lifecycle according to IEC 61511.[2,3] Any gaps that are discovered must be fixed with proper management of change (MoC) procedures.

In the LOPA process, IPLs are rarely challenged, meaning that an IPL operates in low-demand mode (e.g., less than once per year). If an IPL is required to operate more than once per year, then it cannot be considered as an IPL and probability of failure on demand (PFD) credit cannot be claimed. In such scenarios, these IPLs should be considered as initiating events, and the impact consequences of such IPLs must be further analyzed in the LOPA study.

Consider an example for an independent core attribute, an IPL of an SIL-3 rated safety SIF used in the burner management system, to manage the start and shutdown sequence and controls of the entire incinerator burner system. The question is: how to ensure the reliability of this IPL?

The designer of the SIS must select the appropriate instruments, check their reliability using SIL verification calculations, determine the best proof test frequency for each instrument, and verify the SIL certificate submitted by the manufacturer for each device in a particular SIF.

As a result, when choosing such safeguards (e.g., an SIF with SIL-1/2/3) as a credible IPL—which can reduce the SIL target by a magnitude of 1, 2 or 3—the LOPA team should place a greater emphasis on maintaining such an IPL in accordance with IEC 61511's safety lifecycle. The team should clearly document in the LOPA report that such IPLs must adhere to strict enforcement and meet the IEC 61511 functional safety lifecycle phases.

Access security. An access security core characteristic is the use of physical and/or administrative controls with procedures to reduce the possibility of unauthorized changes to the system that could compromise the safety function.

The basic process control system (BPCS) is traditionally used to control process parameters within acceptable limits. Process design and operator actions on alarms play an important role in controlling the processes. Additional layers of protection are available, such as SISs and other mitigation layers like fire and gas, pressure relief valves, dikes and community emergency plans. In any process industry, systems like the BPCS, SIS and FGS form the complete automation system that controls the plant automatically throughout its lifecycle, considering the safety of people, assets and environment.

Technological advancements and increased operational efficiency have enabled most companies to interface operation technology (OT) with enterprise-level information technology (IT) systems, potentially exposing the OT or automation system to cyber threats. Even a robust automation system is vulnerable. These cyber threats, intentional or unintentional, can ultimately result in an abnormal and undesirable shutdown of the plant. Recently, cyber-attacks have caused numerous industrial incidents, so it is essential to study these cyber threats in detail and apply cybersecurity to the automation system.

Apart from automation cybersecurity, various methodolo-

gies are used in the process industries to reduce or prevent the possibility of unauthorized changes to the system, including:

- When changing any parameter of the BPCS and SIS, the security password should be used. Access is limited to those who have been given permission.
- Utilize lock and car seals for valves and/or equipment/devices. This includes the provision of locking the valve or device in the required position.
- Key locking systems are used to ensure that valves or devices are operated in a pre-defined manner and in sequence. A main key is first released from a device or valve from a required position, and then inserted to any remaining devices in the sequence to alter its position as per the predefined sequence documented.
- Use open, close and intermediate position limit switches on isolation or bypass valves that are displaying the actual position to the operator on a human-machine interface (HMI) screen.
- Key lock-type switches should be provided for maintenance override switches and emergency shutdown switches on an operator console to avoid inadvertent operation.
- SIS marshalling panels and system panels should be provided with a unique type of door key and cannot be opened with the door keys of other panels.

During the LOPA study, the team must verify that the facility under consideration has access security systems like those defined here, which must be used under strict administrative controls with proper documentation to be effective. In the LOPA study, such administrative control is examined when a particular safeguard is considered as a credible IPL. Various examples of consequence mitigation systems (CMS) with related risk reduction factors (RRF) and probability of failure on demand are given in IEC 61511-3, Table G7 (PFD).

An interesting example can be seen from that table: an overflow line can be credited as a consequence mitigation system with a risk reduction factor of 100, provided that the overflow line is built to discharge into a containment area that is large enough to handle the hazardous situation. Also, any valves in the line must be administratively regulated for the CMS to be available when needed.

Auditability. Safeguards that can be claimed as an IPL in the LOPA study must be auditable. This requires a robust safety management plan implemented throughout the facility. The audit must be performed to confirm whether the design, operation, maintenance and testing are being implemented as per the guidelines and the procedures mentioned in the functional safety standards. An audit can also help measure the effectiveness of the system and procedure implementation and can be corrected for any gaps. In process industries, the LOPA process is also audited to verify its contents and alignment with the advancement in technologies, reliability data and standards. In this way, the entire LOPA process can be re-validated.

As a result, the team should refer to the Safety Plan document for a specific project during the LOPA study to see whether the requirement for an audit process is specified or not. If an audit requirement is listed, the team should review the content for appropriateness—if not, the team should make a recommendation.

Considering human action as an IPL. Opinions vary when considering human action as a credible IPL in a LOPA study. This includes operator action in response to an alarm within the required time duration, which may depend on various factors like operator experience, operator alertness in response to the alarm, an operator's positive state of mind, documented procedures and the number of multiple tasks being performed by the operator at a time. The overall effectiveness of human action is often less reliable than the automatic actions performed by other means. However, it is too conservative not to consider well-defined human action as a credible IPL.

In the LOPA analysis, human response to an alarm should be considered a reliable IPL. During an actual plant procedure, the operator must monitor many important process parameters to ensure that the process runs smoothly. When all process parameters are normal and within reasonable limits, the operator can relax and observe the process. However, if one of the process parameters reaches the permissible control limits, the operator is under pressure to correct the situation, including conducting comprehensive research. The BPCS system begins sending warnings and alarms to the operator, increasing the difficulty for the operator to control an incident. The operator must take the required measures to ensure that the process returns to its normal operating parameters and that an unwanted shutdown is avoided.

As a result, the operator's function and actions become crucial to restore the process to a safe state. When a human action is considered a reliable ILP in a LOPA analysis, the operator's actions must be carefully observed, evaluated and recorded. That operator's knowledge and awareness should be shared through a robust mechanism such as theoretical training and realistic demonstration through simulation softwares.

PFD values for IPLs. Once the safeguards have been properly analyzed and identified as credible IPLs, the next question is how much credit should be considered as risk reduction measures. The international standard IEC 61511-3 provides examples for the PFD values to be considered for IPLs. However, the final values to be considered in the LOPA study will depend on each organization and its tolerable risk criteria. This will further vary from one process plant to another, as each process has its unique hazards and consequences. Some industrial data are available to consider the PFD value for IPLs, but these should be verified and documented by an organization before conducting the HAZOP/LOPA study.

PFD value data are available for IPLs, documented in literature from industry experience and good engineering practices. For example, one such source of literature is from the Center for Chemical Process Safety (CCPS)[4], which has documented the ranges for PFD values of IPLs used in different companies. For example, the passive IPL like Dike can have PFD values ranging from 1×10^{-2} to 1×10^{-3}; another example would be an active IPL relief valve with PFD values that vary from 1×10^{-1} to 1×10^{-5}; and a BPCS as an IPL can claim the PFD credit ranging from 1×10^{-1} to 1×10^{-2}.

Safeguards are not considered as an IPL. The core attributes mentioned must be verified to qualify safeguards as IPLs—not every safeguard can qualify as an IPL, but every IPL serves as a safeguard. It is essential that the LOPA team analyze each safeguard in detail against the core attributes explained here and decide which safeguards will be credited as an IPL in the LOPA study, as well as the risk reduction magnitude. It has always been difficult for LOPA teams to determine a credible IPL, so IPLs can always be challenged. However, the CCPS has listed some of the safeguards that may not be considered a credible IPL:

- Training and certification
- Procedure (SOP, maintenance procedures, etc.)
- Inspection and testing (FAT, SAT, etc.)
- Any kind of maintenance work
- Display signs and communication
- Fire and gas systems.

Management of Change (MoC). Changes can occur at any stage during a plant's lifecycle (i.e., during the design stage, startup/commissioning or even once the plant is up and running). It is critical that changes be captured at every point and carried out in accordance with a strict MoC protocol. It is possible that new hazards introduced by the changes will affect the current system's integrity, necessitating a new SIL assignment for the existing system.

The functional safety standards recommend that any modifications or changes in the design are properly evaluated, examined, tested and validated before being put into operation. Any additional risk due to these changes or modifications must be re-evaluated and proper risk reduction measures will be applied. Such activities and their documentation will be performed in accordance with proper MoC procedures.

Takeaway. In a particular overpressure scenario, the LOPA team can accept the pressure relief valve (PRV) as an IPL and credit that PRV in the SIL analysis to arrive at the remaining risk reduction magnitude. Due to the time constraints of a LOPA study, the LOPA team generally does not check the actual design of the PRV (e.g., whether the PRV is designed for all required scenarios, selected as per the design basis, designed as per required flow at scenario conditions, installation details, inspection frequency, and whether testing and maintenance are carried out properly). Rather, the LOPA team only ensures that a PRV is available on the P&ID for the particular scenario in question.

Once the LOPA session is over, the LOPA SIL target result is usually not revalidated or challenged until and unless it will create any major issue in design at later stage. The purpose of this example is to emphasize the importance of the selection of a proper IPL in the LOPA study. Because the selected IPL contributes to major risk reduction in the LOPA study, it must be properly analyzed and validated for its intended purpose. In this example of a PRV as a credible IPL, it is validated by auditing the documentation proof for the PRV against requirements like sizing criteria, design basis, installation details, inspection and testing report, maintenance, proof testing procedures, etc.

During the LOPA session for SIL determination, the LOPA team should use data developed from the HAZOP study and follow the LOPA worksheet for further analysis and evaluation, exploring all possible situations or conditions and considering core attributes like independence, functionality, integrity, reliability, access security and auditability before selecting the safeguard as an IPL. If the safeguard under analysis qualifies the relevant and required core attributes, it should be considered as a credible IPL in the LOPA study. HP

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Process Controls, Instrumentation and Automation

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How one analyzer technology can improve a semi-regenerative catalytic reforming process

Going back to the earliest days of the industry, oil refiners have found that breaking down crude oil into marketable fractions does not always result in high volumes of the most profitable products. For example, conventional thermal cracking may create far more naphtha and high-molecular weight fractions than can be sold, while leaving demand for more profitable high-octane gasoline blending components unfilled. As a result, petroleum chemists and process designers have created numerous catalyst-based methods for producing more desirable fractions initially—combined with converting low-value products into higher-value reformates—by using techniques such as:

- Fluid catalytic cracking
- Alkylation
- Continuous-catalyst regenerative reformers
- Semi-regenerative catalytic reforming.

Each process produces some share of high-value products but also creates a stream of byproducts, some more desirable than others. The catalysts tend to be sensitive to feedstock contaminants that in the worst cases, can render the process fully ineffective. Sulfur plays a role here, as it does throughout the whole hydrocarbon chain, as a primary problem for the catalysts and final products, so it must be monitored in all its forms.

This article will concentrate on semi-regenerative catalytic reforming (SRR), which accounts for 60% of reforming capacity worldwide. More specifically, the focus will be on the instrumentation necessary to protect the catalytic action and to optimize the process to produce high-value reformate and desirable byproducts.

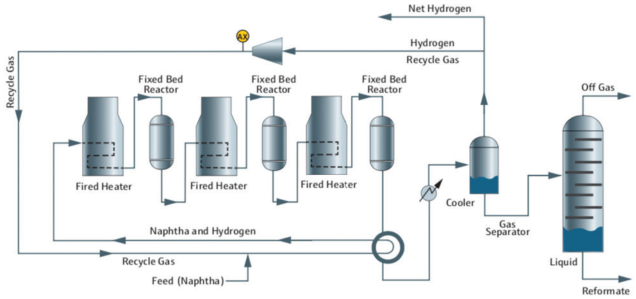

FIG. 1. A typical SRR unit uses a three-stage reaction process. The critical analyzer (AX) is normally placed downstream from the hydrogen recycle gas compressor.

Catalytic reforming process. SRR is a costly process. However, when it is working well, it produces three product streams made from surplus naphtha, increasing overall value well beyond the processing costs.

An SRR unit (**FIG. 1**) converts depentanized naphtha into high-octane aromatic reformates (benzene, toluene and xylenes) used in gasoline blending, as well as a second stream consisting of hydrogen that can be used in other processes. The offgas produced in the fractionation column is then sent for further processing to separate LPG from light hydrocarbons.

An SRR unit has three fixed-bed catalytic reactors, each employing a platinum/rhenium (Pt/Re) catalyst on a chloride alumina support. Fresh naphtha feedstock is mixed with hydrogen and then fed into the reactor train.

Each reactor works at a different stage in the process, breaking and rearranging specific types of molecules in a sequence:

- Reactor 1 handles dehydrogenation and conversion of naphthenes to aromatics
- Reactor 2 isomerizes normal paraffins to isoparaffins
- Reactor 3 facilitates hydrocracking of paraffins.

After passing through the reactors, the stream is cooled to condense the reformate and separate it from the hydrogen. A portion of the hydrogen is recycled and mixed with fresh naphtha feedstock, but much of it is surplus and can be drained off as its own product stream. Raw liquid reformate goes through one more separation step to remove the offgas before use in gasoline blending.

Over time, catalyst efficiency declines as coke builds up in the reactors, so the process must be shut down periodically to burn coke off and restore efficiency. Operational periods between this regeneration processes can be 6 mos–24 mos. After three to four regenerations, the Pt/Re catalyst must be replaced. A full regeneration cycle takes 1 wk–2 wk, which represents a major loss of production capacity. Consequently, careful process control and feedstock monitoring are necessary to minimize the need for maintenance and maximize uptime.

Improved operation begins with careful monitoring. The chemical composition of the hydrogen recycle gas is critical to equipment and process efficiency. There are two main areas of concern: sulfur and chloride.

Hydrogen sulfide (H_2S) and other sulfur compounds (carbonyl sulfide, dimethyl sulfide, tetrahydrothiophene and various mercaptans) poison the Pt/Re catalyst and increase coking, decreasing hydrogen production and reformate yield. Tolerable amounts can be as high as 1 ppm, but if the naphtha feedstock carries sulfur compounds more than this amount, it must pass through a hydrotreater unit to bring these contaminant levels down. Recycled hydrogen must be monitored for sulfur content to maintain process efficiency and minimize regeneration cycles.

The process requires a small amount of H_2O and chloride so the catalyst can perform hydrocracking, isomerization and cyclization conversion reactions efficiently. However, too much H_2O allows formation of corrosive hydrochloric acid (HCl) capable of reducing reformate yield and damaging the equipment. For example, HCl reacts with any traces of ammonia, creating ammonium chloride that can form deposits in the hydrogen recycle compressor. Entrained HCl can also cause corrosion farther downstream in the gas separator heat exchangers. If any equipment requires repairs, the entire SRR unit must be shut down, causing the loss of reformate production. This incurs a hefty financial impact usually well beyond \$1 MM/d in addition to the repair costs.

Tracking the H_2O level in recycled hydrogen gas can determine when it is necessary to reduce the H_2O level and dry down the catalyst. The most effective and practical place to monitor and measure H_2S and H_2O is at the hydrogen recycle return line, downstream of the compressor (**FIG. 1**). Finding the best way to measure these components has provided many challenges for SRR unit operators.

Monitoring H_2S. A traditional method for measuring H_2S is lead-acetate sensing tape, with a length of paper tape impregnated with lead acetate inserted into the recycle hydrogen stream. Over time, it reacts with H_2S to form lead sulfide, which appears as a black solid. The amount formed over a specific period can be measured by an analyzer to indicate the H_2S concentration. The analyzer automatically advances the tape mechanically to expose a fresh section to the gas stream for the prescribed interval. The tape is a consumable item, and the presence of lead acetate (CAS 6080-56-4) and lead sulfide on the spent tape makes it a hazardous waste according to U.S. and EU regulations under the Resource Conservation and Recovery Act (RCRA) Code D002/D003 and EU 16 05 06, respectively, necessitating regulated disposal. This method of analysis can be accurate and require little or no calibration, but it is maintenance intensive and requires special disposal methods.

A more sophisticated approach for measuring sulfur and its compounds is a flame photometric detector analyzer. A sample of the recycled hydrogen is mixed with air and a flammable carrier gas and then burned as a reducing flame in an airless oven. This creates chemiluminescence caused by excited sulfur, which can be measured by a photomultiplier tube and flame photometric detector. This approach can be very accurate, and it is capable of speciating a range of sulfur compounds. However, it requires specialized gas consumables and tends to be maintenance intensive. It is also overkill for this application where a simple H_2S measurement is sufficient for process control.

Tracking the H_2O level in recycled hydrogen gas can determine when it is necessary to reduce the H_2O level and dry down the catalyst.

Monitoring water. Measuring H_2O in the stream calls for entirely different technologies than those mentioned for H_2S, usually necessitating a second analyzer. Traditional H_2O sensors pass a sample of the recycle hydrogen over a surface chilled to well below the dewpoint, causing water vapor to condense on the surface.

A common element of these measurement approaches is that the sensing surface must be heated and dried after each reading to clear the deposit and then rechilled for the next measurement cycle. Aluminum oxide probes can measure moisture without being chilled but must still be heated to take a new reading. This slows down the cycle time, which may result in delayed detection of a changing situation. In addition, some contaminants, if present in the gas stream, can also condense and leave traces on the sensor. These traces tend to build up over time, decreasing measurement accuracy.

TABLE 1. A TDLAS analyzer is designed for hydrocarbon analysis of H_2S (upper table) and H_2O (lower table)

Application data

H_2S analysis	
Target component (analyte)	H_2S in semi-regenerative reformer hydrogen recycled gas
Typical measurement ranges	0 ppm–50 ppm through 0 ppm–300 ppm
Typical repeatability	± 2% of full scale
Measurement response time	1 sec–60 sec
Principle measurement	Differential TDLAS (H_2S scrubber included)
Validation	Certified blend of H_2S in nitrogen balance
H_2O	
Target component (analyte)	H_2O in semi-regenerative reformer hydrogen recycled gas
Typical measurement ranges	0 ppm–50 ppm (control) and 50 ppm–500 ppm (trend)
Typical repeatability	± 1 ppm (control) and ± 10% of reading (trend)
Measurement response time	1 sec–60 sec
Principle measurement	Non-differential TDLAS
Validation	Certified blend of H_2O in pure nitrogen or integrated permeation system

Note: Analysis of H_2S and H_2O simultaneously requires optional accessory

TDLAS: One analyzer for both measurements. Tunable-diode laser absorption spectroscopy (TDLAS) takes an entirely different approach to gas analysis than the technologies just discussed. It applies the characteristics of gases to absorb specific wavelengths of light depending on their chemical composition. How this works has been summarized by the Beer-Lambert law, and it has been applied in numerous ways for industrial gas detection and analysis. TDLAS analyzers are particularly well suited to the components found in natural gas and refining applications, so they are seeing wider deployments within the larger hydrocarbon industry.

For this specific application, the characteristics of H_2S and H_2O are both well understood. Using wavelengths between 1.8 µm–2.8 µm, it is possible to identify absorption peaks in the recycle hydrogen stream caused by H_2S and H_2O, such that they can be quantified simultaneously but separately. For example, the peak caused by H_2O is easy to measure, providing an accurate value of water content in the stream.

A TDLAS sensor (**FIG. 2**) is very simple and has no moving parts outside of valves for the sample handling lines. The gas sample flows into a tube driven by line pressure. Placing the assembly in a heated enclosure ensures that the sample remains fully in a gaseous state, preventing condensation of any components. The sample's temperature and pressure are both monitored for the final measurement calculations.

At one end of the tube is a mirror, and at the other end is a sapphire glass window, behind which is a diode laser next to a detector. The laser puts out a brief flash, sending its beam into the tube where it is reflected to the detector by the mirror, effectively doubling the travel distance through the sample gas.

Software scans the detector to check the most critical wavelengths absorbed by H_2O and H_2S. The algorithm performs its calculations, resulting in the final concentration values (**TABLE 1**), and the sample can be exhausted from the tube. This process happens in a matter of seconds, so measurement intervals can be performed often to suit the requirements of the application.

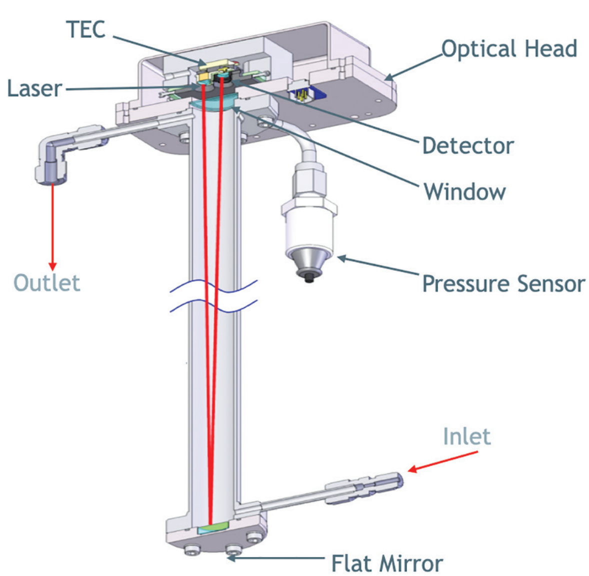

FIG. 2. A TDLAS sensor measures absorption of specific wavelengths of light caused by the presence of specific chemical components.

Since the characteristics of the diode laser and detector are very stable and the absorption characteristics of the subject gases do not change, a TDLAS analyzer exhibits virtually no drift and generally needs no calibration. With no moving parts, gas bottles or delicate mechanisms, TDLAS analyzers are easy to deploy and maintain in the field.

Financial effects and considerations. An SRR unit is expensive to operate, but it yields high-value products. Therefore, it is incumbent on operators to fulfill three objectives:

- Optimize the process to produce the highest possible reformate yield and quality
- Extend the time between catalyst regeneration
- Maximize service life of the expensive catalyst.

Clearly, process availability is paramount, since having an SRR unit down for regeneration or worse—catalyst replacement—results in loss of revenue. Since control of both H_2O and H_2S have a major impact on yield, catalyst life and overall efficiency, the analyzer on the hydrogen recycle line plays a critical role in ensuring the success of the SRR unit.

Simultaneously, the analyzer must deliver its value with minimum cost. A TDLAS analyzer is generally not the lowest cost option, at least based on initial purchase price; however, its reduced operating costs quickly overcome the higher initial investment. There are no consumables, and these devices can run for years without requiring maintenance, recalibration or replacement. Therefore, costs such as analyzer technicians, repair or replacement of probes, as well as an inventory of spare sensor heads are eliminated. The aggregate savings on such tangible costs can easily amount to tens of thousands of dollars per year for each analyzer.

There are also operational considerations. During an abnormal process event, a liquid slug in a sample line can render conventional moisture sensors inoperative for hours or days. For this reason, many refineries have resorted to redundant analyzers, with associated sample handling conditioning and stream switching, which can double or triple the installed costs. These installation complexity costs, combined with consumables and maintenance, quickly negate lower initial purchase costs. Moreover, the revenue gained from even a single additional day of production, or an improvement in reformate quality resulting from tighter process control, greatly exceed the total cost of any analyzer deployment.

TDLAS analyzer technologies for H_2S and H_2O measurements in hydrocarbon processing applications have proven highly accurate and reliable over the long term, even in hostile operating environments. The results of higher unit availability with increased production revenues, combined with reduced operating and maintenance costs, should clarify the decision-making process when selecting an analyzer technology. HP

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LEE NICHOLS, Editor-in-Chief/Associate Publisher

Valves market research analysis details industry leaders in valves manufacturing

Valves are some of the most widely used equipment in hydrocarbon processing facilities. They are imperative to control the flow of natural gas and liquids through processing complexes. According to several industry reports, the global industrial valve market is forecast to eclipse $79 B by 2025 and could increase to more than $90 B by 2030.

Depending on the application, valves come in all shapes and sizes. These include multi-turn, linear motion, quarter-turn, rotary, self-actuated, control and specialty valves. Each type of valve has its own function to ensure intended operations for flow control, and companies around the world are investing in research and development to optimize the usage of valves in industrial applications.

Within the hydrocarbon processing industry, there are countless numbers of valve manufacturers. Which are the most recognizable? Which companies have the best reputation in the industry? What do valve end users/purchasers look for when selecting a valve(s) for their facility's operations? These questions are addressed in *Hydrocarbon Processing's* Valves Market Research Survey report, available at store.gulfenergyinfo.com.

Valves market survey. Earlier this year, *Hydrocarbon Processing* conducted a global survey on the valves market and valve manufacturers. The survey specifically targeted *Hydrocarbon Processing* readers that work within the valves manufacturing, operations, procurement, purchasing, maintenance and/or research and development segments. The survey respondents included personnel from the refining, petrochemicals, gas processing/LNG and pipelines industries. Nearly 70% of respondents either initiate, recommend, specify, approve or purchase valves for their company.

According to the report, most valve procurers value quality and reliability when selecting a valve supplier. These were followed by customer service and innovative technology (**FIG. 1**). Prior to selecting a valve, purchasers will conduct research on a particular valve's operation and usefulness. According to survey respondents, most valve purchasers will go to the supplier's website for additional information on a specific valve. This includes reviewing case studies, whitepapers and any type of catalog/brochure on the valve's operation, technology and specifications.

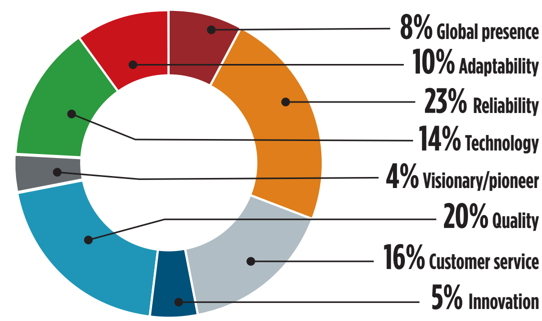

FIG. 1. Attributes valve purchasers value most when selecting a supplier of valves.

The Valves Market Research Survey report also provides a detailed analysis on how valve manufacturers are regarded within the industry. These attributes include rankings on characteristics valued within the industry, including service/technical support, product quality, price, industry/application experience, delivery/lead time, implementation and reputation, among others. The report details these rankings on nearly 40 valve manufacturers around the world.

Globally, survey respondents indicated the following three companies that come to mind first when looking into purchasing valves: Emerson, Flowserve and Crane. These companies were followed by Cameron, Velan and several others.

The full report. The Valves Market Research Survey report provides detailed information on how valve manufacturers are viewed within the energy industry, as well as the key attributes valve purchasers look for when purchasing valves. For more information on the report, contact JNette.Davis-Nichols@GulfEnergyInfo.com. HP

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Aging pressure relief valves: Are you managing history well?

The risks associated with the malfunction or blockage of pressure safety valves (PSVs) are significant, and operating facilities invest heavily in preventive maintenance (PM) programs to mitigate the risks and ensure that PSVs are functioning exactly as per design requirements. Significant resources are allocated to isolate, remove and transport PSVs to a workshop so that they can be returned to service as soon as possible after inspection, testing and overhauling. All field activities are controlled under strict PM programs; however, one thing that is sometimes overlooked is the preparation of traceable historical records, which is important to continuously optimize and improve the program.

Troubleshooting and finding solutions to some notable challenges of PSV data management will not only increase confidence in maintaining the integrity of PSVs, but will also help to reduce PM, saving cost and efforts associated with their testing program.

Collecting clean and reliable data. During root cause analysis (RCA) and frequency review exercises, important observations and findings are often either overlooked or not recorded properly. A report of a pressure vessel, piping or a storage tank handling a non-critical service is usually comprehensive (supported by checklists and pictures) and approved by the most-experienced site inspectors; however, the report of a vacuum relief valve installed on top of the same tank may lack important observations and might have been signed off by someone less experienced in the group with insufficient training.

Therefore, it is vital to establish a robust mechanism onsite to ensure that clean and reliable data is obtained from inspection reports of PSVs. Regular audits of records and proper training of workers can help keep the system alive.

Extracting data from checklists and forms. Large amounts of data kept in printed or hand-written forms require significant time and effort later to visualize the data and observe trends. Rather than writing reports by hand, it is preferred that all reports and checklists related to PSVs are filed directly into the site digital database by each interface involved, and printed certificates or reports are generated through that particular database itself, if required. This will assist later in retrieving, filtering and analyzing the data frequently and in making strong and accurate business decisions based upon technical inputs to reduce cost and risk.

Some sites manage such records in centralized MS Access or Excel data bases, while others are using different inspection data management software (IDMS) for this purpose. Other maintenance applications like Maximo or SAP may also have such databases designed for PSVs.

Inaccurate design and process data. Maintaining accurate and updated process and design data (FIG. 1) of all assets—particularly PSVs—is the foremost requirement of any process safety management (PSM) system. A centralized database link should be provided to all involved in the PM of safety valves so that information can be accessed readily.

This information should be always kept updated. To ensure this is done, the inspector for each PM incident should be held responsible to verify this design data and note remarks for action, if any. If records cannot be updated immediately, then such recommendations should be tracked by IDMS or any other suitable tracking mechanism used at the site.

Missing operating history and vendor data. As per API-576 Section 6.2.4, the operating history of each PSV since its last inspection should be obtained and should include pertinent information. The following updated records are important for proper PM execution and to conduct studies related to reliability or process safety:

- Vendor/manufacturer maintenance manuals
- Spring range catalogues
- Information on upsets and their

Sr. Number	Identification Number	Manufacturer	Type of PSV	Seat/Disc Material	Spring Number	Spring Range	Spring Material	Bellow Material	Construction Date	Set Pressure (psig)	Cold differential Set Pressure (CDTP) (psig)	Back Pressure (Psig)	Service	Criticality	Operating Pressure	Operating Temperature	Link to spec sheet	PM Type (Shutdown /On-stream)	Size Inlet X orifice X outlet	Location Installed
01	15-PSV-XXXX	XXX. inc	Bellow	SS316	G012	0-150	CS	Inconel	1982	1000	1050	50	powder	A	800	350	link	Shut down	2D6	H-1215
02	155-PSV-XXXX																			

FIG. 1. Sample process design data table for PRDs.

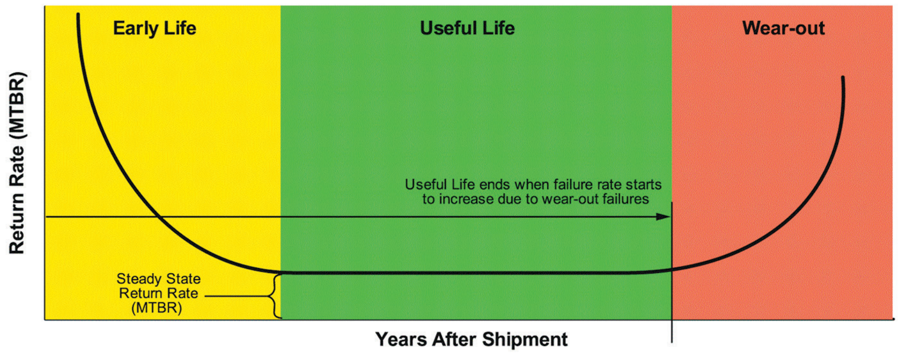

FIG. 2. Typical bathtub curve for PSV internal component failures.

effect on the valve

- The extent of any leakage while in service
- Any other evidence of malfunctioning
- Whether any rupture disks under the PSV have been replaced
- Management of Change (MoC) records and current status of actions
- Record of any frequency review exercises (increase/decrease), along with justification
- RCA related to PSVs
- Drawings of pressure relief devices (PRDs)
- Historical records of failure/popped or leaked in field
- Inspection recommendations record/status
- Test certificates
- Repair record/parts replacement record
- Spare parts list.

Initial visual inspection. Many types of deposits or corrosion products in a PSV may be loose and may drop out during transportation of the valve to the shop for inspection, testing, maintenance and resetting. As soon as a valve is removed from the system, a visual inspection should be made. Visual inspection of a PSV immediately after removal from service provides important information about operating conditions. To minimize errors in the testing and handling of a PSV, each PSV should carry at least two identification marks to show its company equipment, such as identifying tag, stencil on flanges, plate or other means.

As per API-576, "Inlet and outlet piping should also be inspected for the presence of internal deposits, and records should be kept of their condition and cleaning. If necessary, piping should be radiographed or dismantled for inspection and any cleaning to be performed." The type of deposits, degree of restriction due to deposits, and the reason for PSV removal should be recorded.

If such data is recorded properly with evidence, it can help significantly in adjusting accurate inspection intervals.

As-received inspection of PSV in workshop. After a PSV is removed from service, a visual inspection is carried out before cleaning and then it is usually taken to the shop for inspection and repair. An important phase of maintenance is testing to determine the set pressure and tightness of the valve. This testing is usually performed on a test block in facilities for applying pressure to a valve and indicating the pressure applied. This activity is termed as "as-received" inspection.

It has been observed at multiple sites that experienced inspectors are involved only in the final test of the PSV, while as-received inspection activity is left for valve shop technicians who are not fully aware or trained. A qualified inspector must attend the as-received inspection event—this is the heart of an efficient and cost-effective PM program. This inspection activity provides clues about the actual PSV performance during operation and suggests what actions or steps should be taken to improve its performance in the future.

Intervals between PSV testing should be determined by the performance of the valves in the particular service. If the as-received test results are erratic or vary significantly from the cold differential test pressure (CDTP), the inspection interval should be decreased or suitable modifications made to improve the performance. If the PSV is dismantled, then all internals should be checked as per vendor guidelines. PSV internal component failures follow the typical bathtub curve pattern shown in **FIG. 2.** A sudden increase in failure rates can help the inspector determine the equipment's useful life.

The precise recording and reporting format in a PSV testing program depends on each individual company's methodologies; however, as a guideline, key points mentioned in **FIG. 2** and API-576 should be recorded clearly in the test certificate.

Final testing, test results review and record update. Before sending the PSV to the plant or warehouse, a final test should be conducted. Refer to API-576 for key points to be recorded during this final test.

The improper shipment and transportation of PSVs can have detrimental effects on device operation. Refer to API-576 for detailed guidelines regarding transportation.

Once inspections are completed, the results should be updated in the site inspection management software and approved by the inspection supervisor. Any pending actions should be recorded and tracked until implementation. Risk assessment should be carried out by a team if critical recommendations cannot be implemented immediately. The duties and responsibilities for all involved in an inspection and testing program for PSVs should be clearly defined.

Takeaway. The inspection of PSVs provides data that can be evaluated to determine a safe and economical frequency of scheduled inspections. Historical records reflecting periodic test results and service experiences for each relief device should be recorded in full details, as these are valuable leading indicators of asset integrity. HP

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Mitigation actions to prevent vibration during factory tests of SAGD boiler feedwater pumps

The most vital information on the condition of boiler feedwater (BFW) pumps is vibration. The vibration data will determine the mechanical condition of the pump and is an index for acceptance or rejection of the pump during shop performance tests or mechanical running tests at the factory or in the field. This information is also very useful in predictive or preventive maintenance and reliability programs.

This article reviews practical engineering mitigation actions to minimize the risk of vibration of BFW pumps during factory performance or mechanical running tests.

Introduction. BFW pumps are the most critical pumps in many industries, such as power, mining, steam-assisted gravity drainage (SAGD), refining, chemical and petrochemical production, and gas processing. The reliability of BFW pumps has a significant impact on production. **FIG. 1** shows a typical steam generation process in an SAGD plant. The BFW pumps are used to increase the hot water pressure before entering the once-through steam generators (OTSG).

Vibration monitoring of high-pressure BFW pumps is similar to other types of high-speed turbomachinery that employ fluid film bearings. American Petroleum Institute (API) Standard 670[1] covers the details of permanent vibration monitoring systems for critical rotating equipment. High-pressure and high-temperature BFW pumps used in SAGD and many other industries will often be supplied with vibration monitoring systems in accordance with API 670.

The vibration monitoring elements that are normally used for high-pressure and high-temperature BFW pumps are summarized in **TABLE 1.**

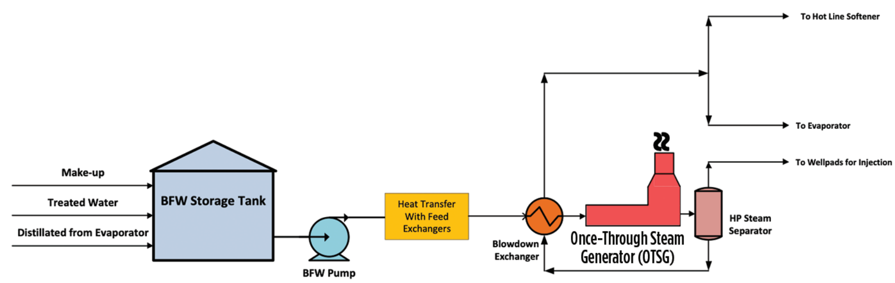

FIG. 1. Typical steam generation process in a SAGD plant.

TABLE 1. Vibration monitoring for high-pressure and high-temperature BFW pumps[1]

Measurement	Description	Role	Signal
Journal bearing	Two proximity probes X-Y (45°) for each journal bearing	Observing relative motion between bearing and shaft	Alarm and shutdown
Thrust bearing	Two proximity probes located at NDE side of thrust bearing	Observing axial position of the shaft within thrust bearing clearance	Alarm and shutdown
Shaft speed/ phase reference	A key phasor transducer to be installed in each independent shaft (non-mechanically coupled); normally a driver shaft to support the driver in uncoupled running condition	To provide a once-per-revolution reference signal for determining vibration phase and shaft rotation speed	
Casing vibration	For machines that are extremely compliant, or as per client specification or vendor's standard		Shutdown
Bearing housing	As per client specification		Shutdown

Vibration consists of four main elements: amplitude, frequency, phase and direction. These elements are the minimum information required to diagnose any vibration from a machine. Vibration measurements of the BFW pump are normally taken by the shaft displacement probes. In addition, API 610 (Clause 6.9.3) requires vibration of the bearing housing to be measured in three planes: vertical, horizontal and axial.[2]

Vibration amplitude in each direction is measured in in./sec or mm/sec (peak). Vibration amplitude is an index of the severity of the problem, and the frequency [in cycles per minute (cpm) or Hz (cycles per sec)] indicates what is causing the vibration. Basically, the frequency of

the vibration peaks helps determine the potential source or cause of the vibration. The vibration phase is the angle of the vibration issue, which is used to determine the dynamic balance and resonance problems (i.e., rotor critical speed).

To determine the cause of vibration, the vibration data can be shown in a wave-form graph, or spectrum. The wave form is a presentation for vibration data in the time domain, and the spectrum is a presentation for the frequency domain. To clarify and simplify the vibration data representation, it is normal practice to use the spectrum, which shows frequency in the horizontal axis and amplitude in the vertical axis. The amplitude can be displacement, velocity or acceleration. The displacement curve is difficult to read at higher frequencies. The velocity curve is the most common curve for the vibration spectrum. The vibration amplitude scales can be linear or logarithmic. By measuring vibration amplitude in three directions and knowing the frequency and pump speed, the cause of pump vibration can be determined by referring to the frequency diagnostic chart.

BFW pumps factory test. **FIG. 2** shows a typical BFW pump performance test setup. According to API 610 (Clause 6.9.3), for pumps with hydrodynamic bearings and proximity probes (e.g., BFW pumps), both vibrations at the bearings housings in three different directions and shaft displacement probes must be measured and meet API 610 allowable vibration limits during a factory performance test.

API 610 requires a factory performance test of the pumps as a mandatory test; however, a mechanical running test is an optional test and is subject to the purchaser's decision. A comparison between a mechanical running test and a performance test of the pumps is summarized in **TABLE 2.**

The proxy probes acceptance criteria in Table 8 of API 610[2] apply only to hydrodynamic journal (sleeve) bearings, which are used for radial loads in BFW pumps. Hydrodynamic tilt-pad type thrust bearings do not have guidelines for acceptance criteria of shaft displacement in API 610. The values for axial movement are normally provided in the instrument alarm setpoints schedule for the pump. Alarm and shutdown values for thrust bearings are identified by the bearing manufacturer.

Low-pressure BFW pumps installed in series at the suction of high-pressure

TABLE 2. Comparison between factory performance test and mechanical running test for BFW pumps as per API 610[2]

		Mechanical running test	
Criterion	**Performance test**	**Bare shaft**	**Combined running test**
Purpose	To confirm that the pump performance and impeller design are based on the original quoted performance curve and to validate that the proposed model is a proven model in terms of characteristics of the impeller	To confirm that pump components work satisfactorily and that all components inside the skid are interacting and working properly together in a combined running test	
Test duration	Not a specific value in API; however, it normally takes about 1 hr	Minimum 4 hr, normally after stabilization of oil and seal temperatures	
Test point	1. Shutoff 2. Minimum continuous stable flow 3. Between 95% and 99% of rated flow 4. Between rated flow and 105% of rated flow 5. Approximately the best efficiency flow 6. End of allowable operating region (**FIG. 3**)	Only between 95% and 99% of rated flow (**FIG. 3**)	
Driver	Test driver—normally a shop motor	Can be test driver or job driver	Job driver
Test components used	Normally pump, job vibration proximity probes, job bearings and job mechanical seals are used in a test stand with a test instruments control system and shop oil system	Similar to performance test	Pump, job driver, job skid, job mechanical seal, job coupling, job instrumentation and control system, job vibration proximity probes, job oil system and seal coolers; however, depends on component availability, and other project constraints may be different
Vibration test	• During the performance test, overall vibration measurements over a range of 5 Hz–1,000 Hz and a Fast Fourier Transform (FFT) spectrum must be made at each test point except shutoff • Vibration measurements shall be made at the following locations: 1. On the bearing housing(s) or equivalent location(s) of all pumps, at the axial, horizontal and vertical positions 2. On the shaft of pumps with hydrodynamic bearings with proximity probes	Normally only with proximity probes	Normally only with proximity probes
Power measurement	By using a torque meter, which has a length higher than the job coupling	Similar to performance test	Power is normally calculated due to limited space for the installation of the torque meter between the pump and driver

pumps for net positive suction head (NPSH) improvement may use anti-friction bearings (i.e., rolling element bearings). The bearing vibration cannot be measured with conventional proximity probes because the rotor is rigidly fixed to the bearing, providing no relative vibratory movement between the shaft and the bearing. For these types of pumps, only bearing housing vibrations need to be measured and compared with the API 610 allowable value for bearing housing during shop or field tests.

API 610 divides the operating flow range of the centrifugal pumps into two regions, as shown in **FIG. 3.** Centrifugal pump vibration varies with flow. The vibration is minimal at the best efficiency flowrate (BEP) and increases as flow is increased or decreased.

At all points of the two regions, the shaft displacement vibration and bearing housings vibration must meet API 610 limits. Sometimes, the vendor interprets Clause 8.4.1 of API 610 for pumps with proximity probes if the bearing housing vibration is out of the API 610 limit; however, if shaft probes are within the limit, then shaft probes data are considered as the basis for accepting or rejecting the pump. This is definitely a misinterpretation of API 610, since both the vibration data from the bearing housing and the proximity probes must meet the API 610 limit. This means that the bearing housing vibrations are measured in units of velocity, and shaft displacement is measured in units of displacement, as shown in Table 8 of API 610.

Normally, displacement is a good measure at lower frequencies (the recommended frequency range is 0 cpm–600 cpm). The failure mode is generally the "stress" caused by the displacement. Velocity measures how often the displacement is applied in a given time period. It is related to the fatigue mode of failure. Velocity amplitude is a good unit of measure in the range of 600 cpm–60,000 cpm. Even at a small displacement amplitude, the repeated motion can cause fatigue failure.

As an example, consider a pump with a vibration due to the vane pass frequency. The vane pass frequency vibration of a centrifugal pump is a hydrodynamically induced vibration at a frequency determined by the number of impeller vanes times the pump cpm. This means that the vibration due to the vane pass frequency has a higher frequency (i.e., greater than 600 cpm). It is less noticeable with shaft displacement probes readings; therefore, most of the vane pass vibration issues of centrifugal pumps are found with bearing housing vibration readings. This means that, during the performance test of a BFW pump, acceptable vibration is measured at 1X, 5X and 7X, according to shaft displacement probes data. Shaft probes may not capture high-frequency vibration over 600 cpm; however, when bearing housing vibration is measured, the pump has vibration at 5X and 7X, which is the vane pass frequency. This is why API 610 requires pumps with hydrodynamic bearings to meet the allowable vibration limits for both the bearing housings vibration and the shaft displacement data.

Causes for pump vibration during factory test. The risk of vibration at flows below and above the BEP is high for BFW pumps with a speed of 3,000 rpm and above during the factory performance test. Qualitative (and quantitative, if required) vibration risk management information from the shop performance test should be provided for new BFW pumps during the bidding stage so that all risks associated with the proposed vendor's model can be recognized and

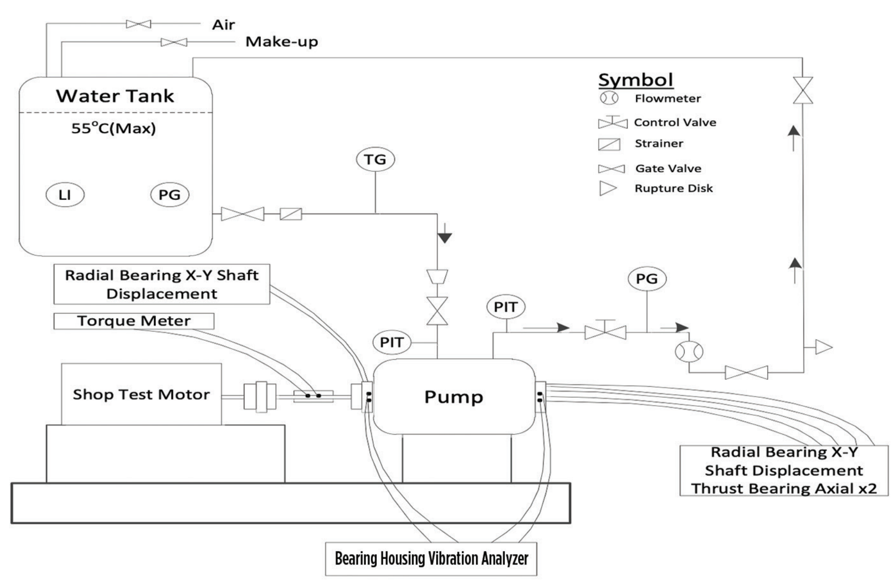

FIG. 2. Typical BFW pump performance test setup.

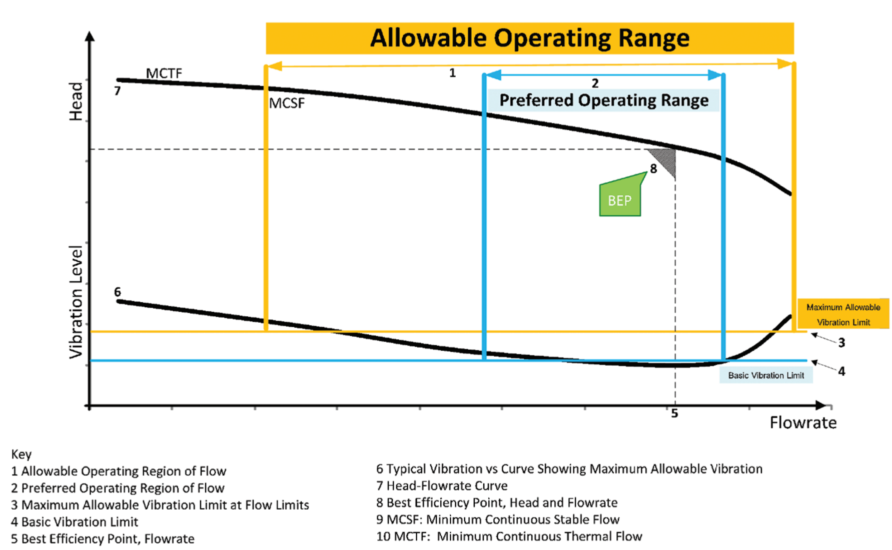

FIG. 3. Regions of operating flow range of the centrifugal pumps based on API 610.[2]

prioritized, and so that adequate resources are allocated to minimize the impact of adverse consequences.

Pump vibration can have many different causes. The main causes of vibration for centrifugal pumps have been reviewed by experts in literature.[4] **FIG. 4** shows the main causes of vibration for BFW pumps in a "fishbone diagram."[4] As **FIG. 4** shows, vibration observed during factory performance or the mechanical running test can be classified into two main categories: mechanical and hydraulic.

The hydraulic forces (i.e., vane pass induced) are significant and normally take place in the BFW pumps. The effect of hydraulic forces may be magnified due to a resonance response within the structure. When the pump is moved from the performance test stand to a job skid for the mechanical running test, the amplitude and composition of the vibration may change, even though no change was made to the pump itself (e.g., impeller trimming).

The conclusion is that the interaction with the mounting structure should be accepted as the root cause, rather than the forces themselves, if the level of bearing housing vibration is still outside of API limits. The hydraulic forces will not be removed by proper installation of equipment and piping. The vendor may refer to the weight of grouting that will be added at the site; however, the increased force of hydraulic pulsations will be absorbed via the bearing housing and pedestals, and not the weight of the skid. If these elements are not rigid, then the effect of the rigid base will be marginal, at best. The hydraulic forces will be dampened by modifications to the impeller and diffuser. Increased clearance between the impeller tip and the volute tongue may help reduce hydraulic pulsations due to the vane pass; however, this may have a negative impact on pump head and efficiency.

Vibration measurements at site test compared to factory test. The vendor may argue that vibration at the site after proper installation will decrease compared to shop measurements. According to ISO 10816-7, 1st Ed., for Category 1 and 2 services, the allowable limit of vibration at the preferred operating region (see "POR" in **FIG. 3**) for the site test is lower than for the shop test, as shown in **TABLE 3.** This may be because ISO makes consideration for temporary foundation and piping during the shop test. However, this is on the basis of presumption that a permanent installation at the site is perfect and free from any design or construction deficiency.

The author believes that this may not be a valid assumption, based on several installations. Under site conditions, vibration levels of BFW pumps are often expected to be higher compared with test bench conditions at the shop due to a few key reasons:

1. Operating temperature for the service at the site is higher than the test water at the factory. Acoustic response will change at operating conditions due to the lower density of liquid compared with the density during the test.
2. Increased hydraulic pulsations at the vane pass and other multiple frequencies due to increased recirculation during turndown may excite structural and acoustic frequencies of the suction and discharge piping. Due to space limitations and other site constraints, it may not be possible to ensure an adequate separation margin with such excitation frequencies and their harmonics. Operating any pump at flowrates outside of the best efficiency will increase hydraulic induced forces due to recirculation. Every effort should be made to avoid resonance of the system during the design stage.
3. Nozzle loads under operating conditions at the site are expected to be higher.

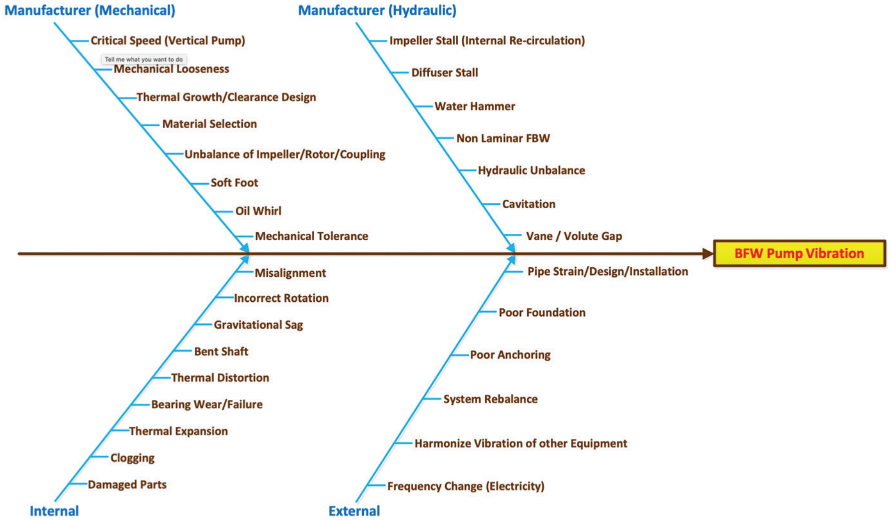

FIG. 4. Main causes of vibration for a BFW pumps in a fishbone diagram.[4]

Practical mitigations for pump vibration during shop performance test. Mitigations of vibration risks during factory test are divided into several main categories, as detailed in the following sections.

Pump and minimum flow sizing. Vane pass hydraulic forces are present in all centrifugal pumps to some degree, and the effects worsen when the pump operates at flowrates outside of the best efficiency. For this reason, the minimum continuous stable flow (MCSF) for pumps must be sized correctly. MCSF is defined as the point where the pump vibration meets the allowable vibration limit.[1] If the pump has a vibration at minimum flow, then the vendor may propose increasing the minimum flow as an acceptable remedy, which will decrease the AOR (**FIG. 3**), and the pump will not be the same product as that originally sold by the vendor.

On the other hand, during the sizing of the BFW pumps, special attention must be given during the early production stage to the size, duty and turndown of the minimum flow recycle valve and cooler, as well as the OTSG quantity, turndown and minimum steam demand. The vendor may say that the pump has an unacceptable vibration at the minimum flow or close to the midpoint, without assuming that the pump will

operate for a long time in these conditions. However, in SAGD operation, it is likely that the BFW pump will operate near the minimum flow or midpoint flow, due to the lesser steam demand for injection to the well.

One BFW pump is common for two OTSGs; therefore, the minimum turndown of the steam generator must be considered in the sizing of the BFW pumps. For example, if the minimum turndown of one steam generator is 50 m^3/hr and the minimum flow of the pump is 100 m^3/hr, then the pump will run at 150 m^3/hr at minimum, which may be far from the BEP. This is because, at reduced capacity, pulsations will increase due to the greater head and turbulence energy. Force response is significant for high-energy pumps such as BFW pumps. Apart from high vibrations, operation under high turndown conditions causes an increase on both radial and axial loads on the bearings.

Pump selection. The ranges for AOR and POR (see **FIG. 3**) indicated on the vendor's preliminary performance curve and data sheet are preliminary and must be validated by the purchaser and owner. The customer should ask the vendor to provide real test and vibration data for the same model in a previous job, as well as for a model of similar size.

The owner and the engineering company should witness the operating condition of the pump after installation and talk to the end users directly about vibration data after site installation and commencement of operation. The vendor manufacturing and test facility should be inspected, as well, since some vendors have only an assembly and test shop and run their design and engineering from another country. Warranty terms should be carefully addressed and consider the risk of unacceptable bearing housing and shaft displacement vibration during factory performance or mechanical running tests.

In addition, measured force response resonance testing should be carried out to determine the system's natural frequencies and dampening characteristics. Impact testing identifies natural frequencies and the resulting mode shapes at various frequencies of the pump. Upon obtaining natural frequency data, a designer can modify structural, pedestal, skid design, etc., to shift the natural frequencies away from the expected operating speeds.

TABLE 3. ISO 10816-7 Vibration allowable level at site and shop for Category 1 and 2 services

	Category 1 (critical application), > 200 kW	Category 2 (pump for less critical or general application), < 200 kW
POR in situ	3.5 mm/sec RMS	4.2 mm/sec RMS
POR in situ	4.4 mm/sec RMS	5.2 mm/sec RMS
POR FAT	4.3 mm/sec RMS	5.2 mm/sec RMS
POR FAT	5 mm/sec RMS	6.1 mm/sec RMS

TABLE 4. Allowable vibration limit based on API 610, ISO 10816-7 and Hydraulic Institute 9.6.4

Region	API 610	ISO 10816-7	Hydraulic Institute 9.6.4
AOR FAT	3.048 mm/sec RMS	4.3 mm/sec RMS	4.3 mm/sec RMS
POR FAT	3.9624 mm/sec RMS	5 mm/sec RMS	5.59 mm/sec RMS

Acceptance criteria. API 610 divides the operating flow range of centrifugal pumps into two regions, as shown in **FIG. 3.** At all points of the two regions, the shaft displacement vibration and bearing housing vibration must meet API 610 limits. If the pump does not meet the API 610 vibration limit, then the vendor may ask to test the pump at the site, as per ISO 10816-7 or the Hydraulic Institute's vibration limit. Note, however, that pumps not in compliance with API 610 are in a major non-conformance situation. **TABLE 4** shows a comparison for vibration limit based on API 610 (Table 7 in the 11th Ed.), ISO 10816-7 (Annex A) and Hydraulic Institute 9.6.4 (for critical application, P > 200 kW).

Pump support pedestal. Pump pedestals should not be tall and should be "full box" design. The full box design may be filled with grout at the site to add mass, if necessary. The vendor may decide to increase the pedestal height due to the skid design, a slope requirement for the lube oil return line, the centerline of motor coupling or another valid reason.

The dimensions of the pedestal should be compared with the dimensions shown for the proposed model in the catalog. No industry standard has been established for pedestal height for BB3 and BB5 centrifugal pumps; height is per manufacturer standard. However, a lower center of gravity for pump assembly is recommended for better rigidity and stability. Therefore, the vendor design should provide a pedestal height that is sufficient to ensure access to all nozzles around the casing, including casing drains. Proper gravity drainage of the return oil line can be achieved by installation of the oil console out of a pump skid in a lower elevation.

Bearing housing pedestal stiffness. It is recommended to increase the stiffness or rigidity of the bearing housing. This can be achieved by a metal-to-metal fit and increased surface contact for the bearing housing flange with a drip pocket flange. Also, additional ribs should be provided for the drip pocket supporting the bearing housing.

Bearing housing material. The weight or material of the bearing housing should be increased by changing the material and/or thickness. Bearing housing made with carbon steel material (e.g., A261 WCB) should be validated for proper vibration measurement data during installation at the site. Materials with better vibration dampening capacity, such as cast iron, should be considered. The vendor should confirm if the option of cast iron construction for the bearing housing is feasible. Also, a study for separate bearing housing design (or, alternatively, bearing housing that is integral with the casing) should be conducted.

Isolation and damping. A dynamic absorber is a device designed to counteract an initial exciting force by creating the same resonate frequency through its own vibratory force. Attaching the dynamic absorber in series with the resonate bearing pedestal creates an anti-node for the bearing pedestal, thereby significantly reducing the resonate vibration amplitude.

Acoustic studies. BFW pumps are ei-

> Ask a competent inspector (preferably API-SIRE certified) with pump knowledge for shop inspection activities of the pumps. Competent means having necessary, valid and relevant training, knowledge and experience. There is a big difference between watching/seeing and inspection.

ther BB3 or BB5 type pumps with 10–12 stages of opposed impellers to help hydraulically balance the thrust force. The internal crossover passage of opposed impeller design is long and has potential for acoustic vibrations. Increased vibrations at the vane pass frequencies of 5X and 7X at the reduced flow between the minimum continuous stable flow and the midpoint are probable; therefore, a critical evaluation for acoustic resonance is recommended at the early design stage.

The measurement record for noise levels will be helpful to see if a correlation exists with cases of increased vibrations. Acoustic phenomena triggered by the vane pass may manifest with an increase in rattling sound.

Test stand. The configuration, layout and support of the temporary test piping is important. The test motor health, alignment and length of the torque meter may influence the pump vibration. During the bidding stage, it is important to verify the vendor's test facility. Sometimes, due to limited space and access at the test area, the test setup may have a flow regulation valve close to the pump discharge flange. Considering high discharge pressure, significant turbulence is expected close to the discharge flange due to throttling for reduced capacity. Turbulence will induce higher vibrations in the pipe and pump casing, thereby increasing vibrations on the bearing housing.

The site configuration will eliminate such a force response if the suction and discharge piping system is carefully designed and installed. The risk for resonance on the test stand should be determined and analyzed. The rigid foundation of the test stand should have a suitable dampening effect compared to the actual foundation.

Nozzles size and location. Suction and discharge nozzle sizes should be checked carefully. The vendor should use the product's hydraulic range chart to demonstrate that the duty points for the proposed pump (with the proposed suction and discharge nozzles size and the double-suction, first-stage impeller) is at the BEP. Locations of the suction and discharge nozzles are also important to mitigate the risk of vibration at the bearing housing. The discharge nozzle of BB3 or BB5 is normally at the middle of the casing with enough distance and symmetry to both inboard (IB) and outboard (OB) bearing housings. However, the suction nozzles normally located near the IB bearing housing have an effect on the vibration of the IB bearing housing.

Operating deflection shape (ODS) analysis. ODS analysis is the simplest way to check and validate how a pump and structure react during operation condition. These shapes are used to determine the effects on a structure of internal and external excitation forces. ODS analysis provides an animated view of a pump and structure's motion at specified frequencies. Data for the model are taken for the structure under typical operating conditions. ODS analysis for similar jobs may provide good supporting documentation at the bidding stage to confirm that the proposed model has a proven design, in terms of vibration.

Balance drum location. The location of the balancing drum is important. The rotor has additional damping due to the balance drum. Other things being equal, lower vibrations can be expected on the bearing housing close to the balance drum. Normally, the balance drum is recommended to be located close to the OB. Shaft vibration readings and bearing housing vibration data for pumps with a balance drum near the OB results in IB vibration being greater than OB vibration, for all operating points.

Recommendations.

Practical mitigation actions to minimize the risk of vibration of BFW pumps during shop performance or mechanical running tests were reviewed.

Pumps with hydrodynamic bearings and proximity probes (e.g., BFW pumps) should have a vibration value at the bearing housing and shaft displacement probes measuring within the allowable limit of API 610 at all points on the performance curve during the factory test. Shaft displacement probes data should not be used as the only basis for acceptance or rejection of the pump during the shop performance test.

An engineer should also ask the vendor for a test experience list and vibration data for the quoted model during the bidding stage. The previous experience of the vendor must be carefully reviewed, validated and witnessed (when necessary), and the engineer should check that both bearing housing vibration and shaft displacement probes data of the previous, similar model were within API 610 allowable vibration limits. HP

ACKNOWLEDGMENTS

The author is very grateful to practitioners, field and factory workers for sharing their experience. Special thanks is also given to Mr. Mehrshad Rahimi for the drafting figures.

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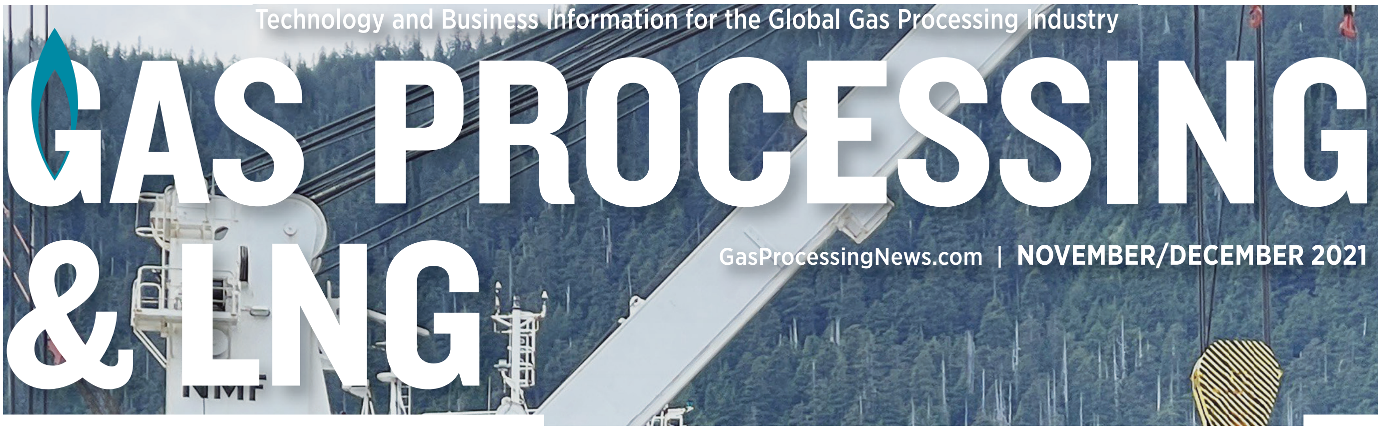

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PLANT DESIGN & PROCESS CONTROL

TREATING & NGL

PLANT OPTIMIZATION

Gulf Energy

EDITORIAL COMMENT

A. BLUME,
Editor-in-Chief

With Henry Hub (HH) natural gas prices almost twice as high year over year, as of the time of writing in late October, and similarly record-high prices in Europe and Asia driving competition for LNG, the gas market is rife with tension. Below-average storage levels heading into the heating season have boosted demand for U.S. LNG and exacerbated tensions among EU governments.

With electricity prices expected to soar over the 2021–2022 winter, the European Commission is floating the idea of joint purchases of natural gas by EU countries and an overhaul of power markets. Wealthy, climate-driven countries in northern Europe insist that the present crisis is not tied to EU green ambitions, while southern and eastern nations are more skeptical. The EU's plan to ban new fossil fuel cars by 2035 and slap a carbon tax on home heating could prove tough to sell to consumers if double-digit power price increases materialize over the winter.

Meanwhile, the U.S. EIA expects HH prices to calm after Q1 2022 as LNG production capacity increases, outpacing demand for LNG export supply. HH prices are anticipated to average \$4.01/MMBtu in 2022, although the unreliability of weather forecasts make the price outlook for this winter uncertain.

Just as disruption in gas markets is sure to bring new direction, this Editorial Comment marks the end of my leadership of *Gas Processing & LNG* after 8.5 years. Be on the lookout for a change in the editorial roster in 2022, and thanks for your many years of insightful feedback and readership! **GP**

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Cover Image: Offloading of the main cryogenic heat exchanger at LNG Canada in Kitimat, British Columbia, Canada. Photo courtesy of Canada LNG.

China signs huge LNG deals with Venture Global

China has signed three large LNG deals with U.S. exporter Venture Global LNG, as the world's second-biggest economy looks to secure long-term supplies amid soaring gas prices and domestic power shortages. The agreements with China's state oil giant, Sinopec, include two 20-yr deals for a combined 4 MMtpy of LNG.

Those deals bring Venture Global's planned, 20-MMtpy plant in Plaquemines, Louisiana one step closer to a final investment decision (FID), which is expected by the end of 2021.

For China, which has this year overtaken Japan as the world's number-one LNG buyer, the deals will be its single-largest LNG trade agreement in terms of volume without an equity stake. The deals will also double the volume of China's imports from the U.S., its sixth-largest supplier in 2020, with volumes at 3.1 MMt.

Both of the 20-yr deals are sales-and-purchase agreements, according to the notice. One is for 2.8 MMtpy of LNG on a free-on-board (FOB) basis, and the other is for 1.2 MMtpy on a delivered-at-place-unloaded (DPU) basis. Venture Global also signed a third deal with Unipec, the trading arm of Sinopec, to supply 1 MMtpy of LNG from its Calcasieu Pass Facility for 3 yr from March 1, 2023.

Major Chinese companies are in advanced talks with U.S. exporters to secure long-term LNG supplies as rocketing gas prices and domestic power shortages heighten concerns about the country's fuel security. In addition, U.S. Henry Hub futures-linked pricing offers a hedge to Chinese buyers that are heavily exposed to Brent-based pricing.

Venture Global is building or developing over 50 MMtpy of LNG production capacity in Louisiana, including two 10-MMtpy phases at Plaquemines, with the first phase expected to enter commercial service in 2024.

Vietnam approves $2.3-B LNG power project

Vietnam has approved a $2.3-B LNG power plant, due to be co-developed by Vietnamese and South Korean companies and to start commercial operations in 2026 or 2027. The plant, which will be based in the central province of Quang Tri, will have an initial capacity of 1,500 MW.

Vietnam's T&T group and South Korea's Korea Southern Power Corp., Hanwha Energy Corp. and Korea Gas Corp. are the project's main developers.

Vietnam's demand for electricity is forecast to rise 10%/yr in the coming years, and its LNG imports are expected to increase to 10 metric MMt by 2030 and to 15 metric MMt by 2035. The country plans to begin importing LNG from 2022.

Israel considers pipeline to boost gas to Egypt

Israel is considering the construction of a new onshore pipeline to Egypt to quickly boost natural gas exports to its neighbor in the wake of the recent tightening of global supplies. The pipeline, which would connect the Israeli and Egyptian natural gas grids through the north of the Sinai peninsula, is estimated to cost around $200 MM and could be operational within 24 mos.

A new onshore pipeline, coupled with plans to construct a second subsea pipeline to Egypt within a few years, would further cement Israel's position as a key energy hub in the eastern Mediterranean. The pipeline, which is in the process of obtaining approval from local authorities, would be owned by Israel Natural Gas Lines.

Israel became a key supplier of natural gas to Egypt in January 2020 after starting production from its Tamar and Leviathan offshore gas fields. Around 5 Bm^{3}y of gas is supplied via a subsea pipeline connecting Israel and the Egyptian Sinai peninsula. The new pipeline would allow increased supplies into Egypt by an additional 3 Bm^{3}y–5 Bm^{3}y. The gas will supply Egypt's power grid and also help boost LNG exports from Egypt to Europe and Asia.

Israel's export capacity is expected to increase to 8 Bm^{3}y by 2023 from the present 5 Bm^{3}y due to the debottlenecking of existing infrastructure and the expansion of the giant Leviathan field, operated by Chevron Corp. and Delek Group. Output will rise further when Energean starts production from the Karish and Tanin gas development, planned for mid-2022.

The onshore pipeline would not impact Israel and Egypt's plan to construct a second subsea pipeline to supply Egypt's Idku and Damietta LNG plants, from which fuel can be reexported to Europe and Asia.

Russia targets year-round shipping via warming Northern Sea

In 2022 or 2023, Russia plans to begin year-round shipping via the Northern Sea Route that passes through the Arctic.

The country has invested heavily in infrastructure to develop the Northern Sea Route and wants it to become a major shipping lane as the Arctic warms at a faster rate than the rest of the world, due to climate change. It is not presently used in winter due to thick ice cover.

The route, which runs along Russia's northern flank, is used to ship hydrocarbons and other resources for up to 9 months per year. Officials want to increase cargo volumes shipped through the route to 80 metric MMtpy. Last year, 33 metric MMt of cargo were shipped on the route.

President Vladimir Putin wants to start regular container shipments through the shipping lane. Russia also plans to build icebreakers powered by LNG, as well as heavy-duty icebreakers to develop the route and make it more suitable for year-round navigation.

EU floats joint gas purchasing plan

Amid high natural gas prices in early autumn, the European Commission outlined measures that the EU could use to combat surging energy prices: joint gas purchasing among countries as a way to cushion price spikes.

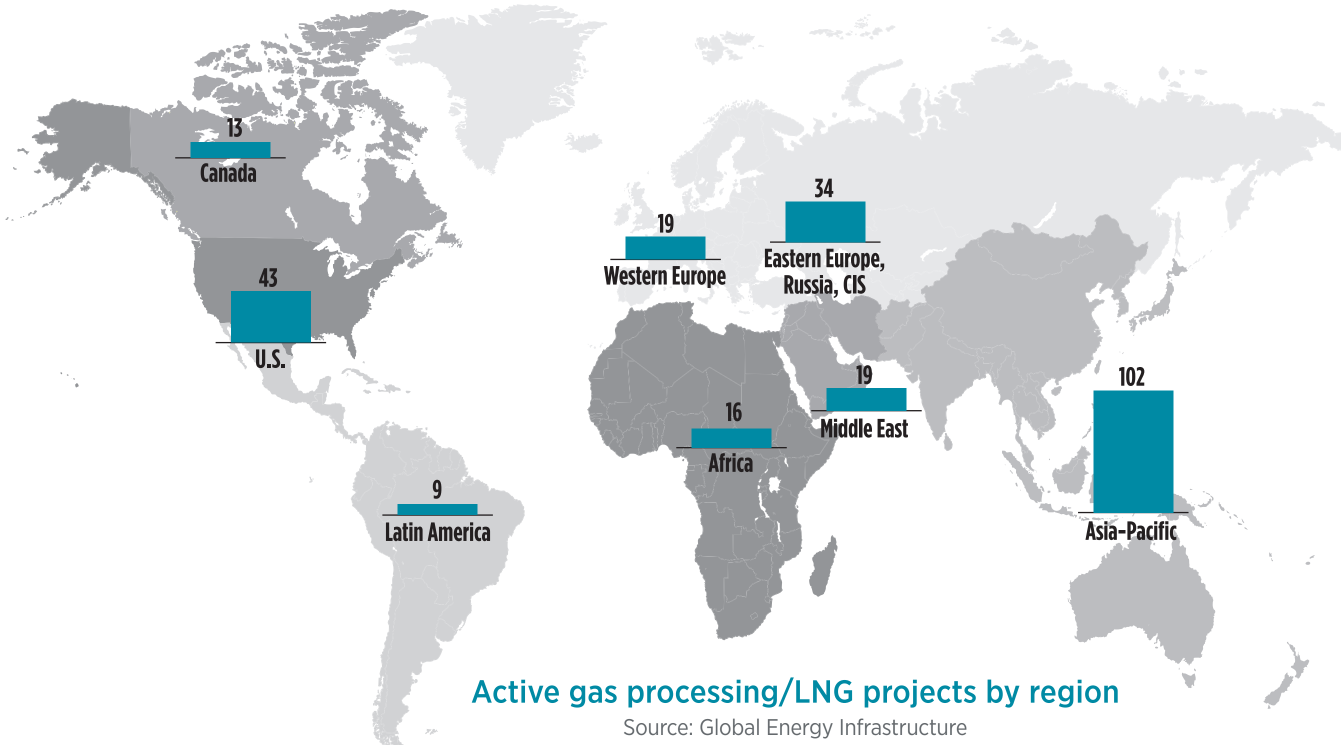
The EC intends to "explore the potential benefits" of member states jointly buying strategic reserves of gas. Countries' participation in such a scheme would be voluntary.

Brussels is also investigating "all allegations of possible anti-competitive commercial conduct" by gas suppliers, following accusations from some EU countries that say Russia's Gazprom has withheld supply to push up European gas prices.

Russian President Vladimir Putin said electric power shortages were behind the October gas price increases in Europe, and that Moscow is willing to discuss additional action, saying that there needed to be agreement on how to balance energy markets.

LEE NICHOLS, Editor-in-Chief/Associate Publisher, *Hydrocarbon Processing*

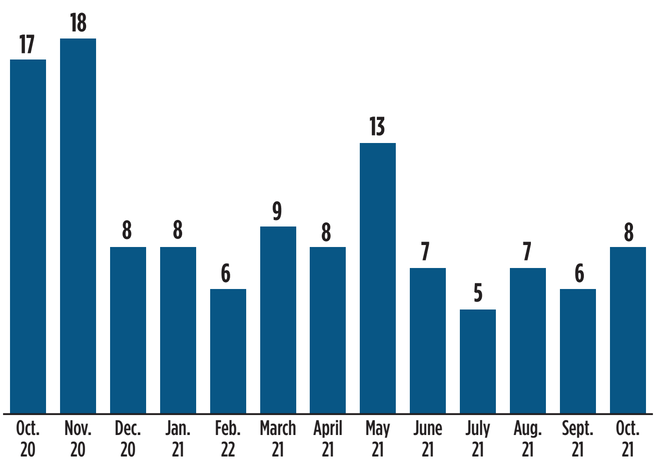
Gulf Energy Information's Global Energy Infrastructure database is tracking more than 250 active gas processing/LNG projects around the world. At 40%, the Asia-Pacific holds the largest market share, followed by the U.S. and Eastern Europe, Russia and the Commonwealth of Independent States. Asian projects are dominated by new LNG regasification projects in China and India. Both nations are investing heavily in new natural gas import infrastructure and pipeline distribution capacity to satisfy increasing demand for industrial use and power generation. The U.S. and Russia continue to significantly increase LNG export capacity to supply developing regions with much-needed natural gas. **GP**



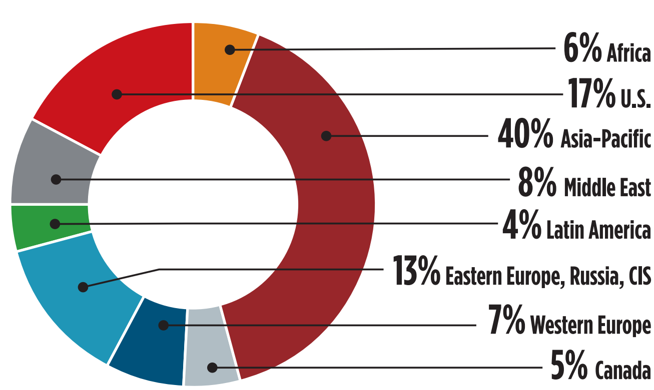
Active gas processing/LNG projects by region

Source: Global Energy Infrastructure

New gas processing/LNG project announcements, October 2020–October 2021



Active gas processing/LNG market share by region



Looking beyond processing efficiency to optimize operations

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When facility owners think about plant optimization, the emphasis often is on maximizing throughput and, in many cases, that means focusing on processing equipment. While production efficiency is essential to profitability, there are other contributors to optimized operations—including reliability, availability and sustainability—that apply to every facet of operations.

Utility piping and fire safety systems, which are essential to safe and continuous operations, can have a considerable impact on plant uptime, illustrating just how critical they are to maintaining productivity.

Scope of impact. Piping for utility and fire safety systems often makes up more than one-third of the total piping in a gas processing facility and includes a broad range of systems from utility air, compressed air and instrument air, to nitrogen, inert gases, lube oils, glycol and water.

A newbuild LNG facility requires all of these utility and safety systems. Taking into account only some of these essential functions, that could amount to thousands of linear feet of pipe, including small-diameter pipe for fire safety systems and large-bore piping systems in excess of 24 in. for water and glycol, inert gas and air. Pipe material can be carbon steel, stainless steel or nonmetallic, in applications that require flexibility or chemical resistance and in operations where wide temperature variations occur.

To date, joining methods for these piping systems have included welding and flanging for steel pipe and heat fusion for HDPE, but the inherent drawbacks of these approaches are significant, and all of them make installation and maintenance costly and time-consuming.

Considering a different approach. Although gas processing facilities have relied on flanging and welding for steel piping in utility systems, these joining methods are not ideal and often are not cost-effective. The shortcomings are well known, but there is a reluctance to move away from a solution that works. Achieving greater efficiencies, however, means working in a different way and recognizing the value of a proven solution.

For more than a century, grooved mechanical joining solutions have been used across pipe materials and sizes to provide a solution that mitigates risk, maximizes productivity and minimizes construction schedule. This proven technology, adopted in a variety of chemical and energy sectors including chemical, petrochemical and LNG, is changing the way piping systems are designed, installed and maintained.

The coupling design is simple and consists of four basic components: grooved pipe, coupling housings, bolts and nuts, and the engineered elastomeric gasket. The gasket engages the full pipe end circumference, fitting securely into the grooves to create a unified joint, which is then enclosed by the coupling housings and bolted on. The housings are installed using simple hand tools, and as the bolts and nuts are tightened, the elastomer gasket creates a triple-seal effect on the pipe ends. The gasket sealing lips press down onto the pipe ends when the system is pressurized, creating a leak-tight joint.

The couplings, which can be installed up to 10 times faster than other joining methods, require only a visual inspection to verify proper makeup. If the coupling needs to be reinstalled, the process only takes a matter of minutes. Couplings are manufactured as rigid or flexible joints for a range of applications and can be installed on nearly any size pipe.

Installing grooved couplings is straightforward because assembly requires only standard hand tools. On a recent project, more than 120 pipefitters were trained to install mechanical grooved couplings in less than 20 minutes and were immediately able to execute the installation. On average, even for these first-time installers, it was possible to install the grooved mechanical system in half the time it would have taken to weld.

Furthermore, the visual inspectability of the mechanical grooved couplings enabled these installers to verify that every joint was properly made up, ensuring system reliability without the additional quality control documentation that would have been required for weld inspection.

Addressing challenges, adding value. The benefits of grooved mechanical couplings are being realized on projects in a range of applications both inside and outside the facility.

In a recent LNG facility expansion, owners encountered a problem transitioning a carbon steel pipeline from aboveground to belowground, where it would be buried in unstable soil. Experts had predicted that over the course of 2 yr, the pipe would experience 5 in.–6 in. of movement caused by settling, which exceeded the performance capability of the previously installed flanged joints. Resolving this problem was simple, using flexible grooved couplings installed in a series at the transition point. The mechanical couplings accommodated the dynamic movement, and because they are easy to install, the joints could be made up and inspected quickly to keep the project on track. This solution maximized onsite productivity throughout installation and will deliver additional gains down the road when inspections need to be performed or maintenance is required.

Grooved couplings reduce maintenance downtime and, with a union at every joint, simplify system access. Couplings are easily disassembled and removed from the joint and can be reassembled as quickly as the initial installation. The de-

sign simplifies maintenance, repair, and future expansions or modifications. When properly installed and operated within design specifications, grooved mechanical couplings are designed to last the life of the system (**FIG. 1**).

Facilitating the move toward net zero. Grooved couplings also help owners drive toward net-zero goals by decreasing emissions and reducing the carbon footprint of operations. They do not generate the byproducts and fumes that are produced by alternative joining solutions. Also, because hand tools can be used for installation, assembling the couplings requires no gas- or electric-powered tools, so no

FIG. 1. Grooved mechanical couplings not only deliver installation and maintenance efficiencies in gas processing and LNG facilities, but also improve site safety and sustainability. Photo courtesy of Victaulic.

CO_2 emissions are introduced by the power source.

These couplings also are proven to reduce fugitive emissions. The gaskets used in proprietary grooved couplings were tested at a third-party laboratory to assess their efficacy compared to standard ANSI Class 600 flanges. High-pressure helium was applied to four coupling assemblies and four flange assemblies and thermally cycled to determine the effects of elevated temperature on emission rates. The couplings and gaskets exceeded the performance of standard flange connections, providing consistent and uniform sealing at both the temperature and pressure extremes of the tests.

Enabling technology changes the status quo. Profitability depends on performance, and that means much more than improving throughput. Capturing efficiencies also means ensuring reliability and system availability and using technologies that enable sustainable operations.

Grooved solutions deliver on every requirement, and successful installations prove that they can improve operations in the gas processing sector. **GP**

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Building the foundation for a low-emission LNG facility

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Electric LNG (i.e., eLNG) has emerged as a powerful decarbonization pathway for LNG project stakeholders. However, the development of these facilities often presents unique technical challenges not encountered with traditional gas turbine-driven designs.

One critical facet that needs to be addressed is integrating the onsite power generation unit (PGU) and the liquefaction island. Traditionally, the parties involved with the design and engineering of these facilities have operated in silos, often with different mindsets and objectives. In addition to potentially limiting opportunities for total plant decarbonization, this type of project approach increases the likelihood of encountering technical problems that can lead to costly rework. It can also present issues if and when plant modifications need to be made or operating conditions change (e.g., capacity or power demand increases, renewable energy sources are integrated, etc.)

To ensure optimum uptime and create the lowest possible plant emissions profile, a more holistic design approach is required where there is close collaboration between the power/utility and liquefaction design teams starting in the early phases of the project timeline. At Siemens Energy, we refer to this concept as "integrated electrical and LNG plant design."[1]

eLNG vs. mechanically driven plants.

In a traditional gas turbine-driven LNG facility, the driver selection determines how much power is available for compressors and dictates LNG train capacity. While the liquefaction plant owner-operator provides a desired nominal design capacity for the facility based on factors such as available feed, LNG demand, export regulations, financial decisions, etc., gas turbines come in discrete sizes. As a result, the LNG capacity is usually designed to fully utilize the site-rated gas turbine power for maximum profit.

When designing an eLNG facility, capacity limitations may appear superfluous in that the operator is no longer constrained by the gas turbine driver when setting the capacity. However, because electricity generation is primarily done with gas turbines, determining the "sweet spot" of the PGU in terms of efficiency is essential for maximizing efficiency and optimizing CAPEX. This sweet spot can be determined by utilizing cogeneration and a close match between the site-rated power and the LNG production (**FIG. 1**).

Moreover, developing the liquefaction and power facilities separately results in a highly inefficient project workflow. In most cases, the process design team starts by creating the electrical demand for the main refrigeration trains and then feeding it to the PGU team. Often, the PGU team must feed information back to the process team, which then initiates additional rework.

Adopting an integrated PGU-LNG train facility development approach makes it possible to avoid this iterative design process and identify issues early on during FEED that can impact efficiency, availability, CAPEX and overall plant emissions.

Production and use of process heat.

With mechanically driven trains, refrigerant compressor driver power and process heat come from the mechanical drive gas turbines. In an eLNG facility, however, the paradigm is shifted. The thermal heating required for the LNG process units comes from the PGU through waste heat extraction or steam. This requires evaluating the needed heat and power during the LNG facility's normal, steady-state operation and during startup, shutdown and upset scenarios.

One example of this is during startup—a topic typically ignored in the early

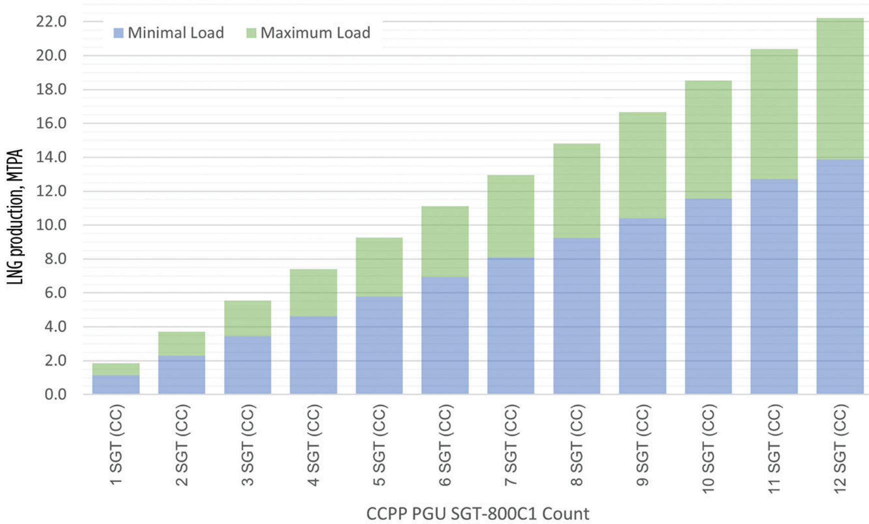

FIG. 1. LNG production rates of various combined-cycle power plants using SGT-800C1 in combination with cogeneration. As the power generation facility increases in capacity, so does the range of LNG production for a given design.

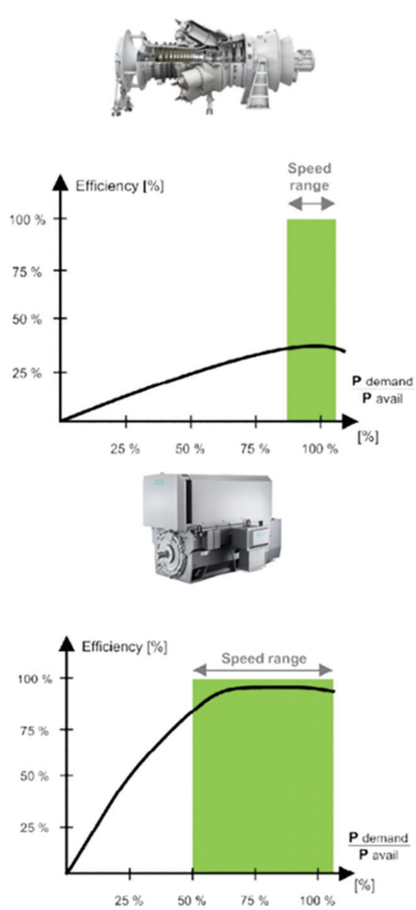

FIG. 2. Refrigerant compressor driver efficiency: Gas turbine (top) vs. electric motor (bottom).

design stages of the LNG train in relation to the PGU. At an eLNG plant with a PGU operating in island mode, a delicate balance exists between the amount of steam available for process heating relative to the electrical load demand. Logically, the LNG train starts up from the upstream gas processing unit and cascades down to the liquefaction and storage facilities. This standard startup sequence results in a higher ratio of process heating load to electrical power demand during the first steps of a startup, compared to the ratio of design to steady-state. Exceeding the design ratio indicates an imbalance to the system that would require supplemental process heat to be supplied by another source.

Adding a furnace to provide the heat necessary during startup is a potential solution. However, this will result in higher emissions and add a piece of equipment typically not used in normal operations. A temporary furnace rental is another solution, but it may not be feasible, depending on the LNG train capacity. In some instances, it may also be possible to address the heat imbalance by proposing an alternate startup sequence.

The best option will ultimately be dictated by the unique requirements of the plant and operator, and it can be determined only by evaluating the power plant and liquefaction island together, holistically.

Electrical grid stability. Equally important to ensuring a reliable supply of process heat is the stability of the plant's electrical grid.

PGUs in LNG facilities are typically classified as operating in island mode. This mode is coined as such because the PGUs are not connected to an external power grid—e.g., a local municipality. If the electrical grid is unstable, then the LNG train would be forced to trip and shut down, resulting in a loss of revenue.

In such installations, maintaining a balance between power generation and electrical load consumed is crucial. Active and reactive power imbalance can lead to frequency and voltage deviations, and without sufficient reserve, frequency and voltage regulation may present serious problems. Therefore, the electrical system must ensure voltage and frequency stability in all scenarios at all times. A comprehensive electrical system evaluation is essential for ensuring reliability at any continuous process facility; however, it is particularly so when renewables enter the energy mix.

Several LNG projects across the world are now evaluating the use of renewable power sources to reduce emissions. More specifically, this is being considered in combination with conventional natural gas generation to create a hybrid system.

Integration of renewables adds benefits, but it also complicates the stabilization of the plant electrical grid. For example, a photovoltaic (PV) plant can provide support during over-frequency cases. The challenge, then, is making the energy system flexible enough to cope with this transition.

The renewable power output from solar or wind is highly dependent on renewable source availability. A possible solution to balance this intermittence is to use energy storage. Power production and demand can be balanced by storing electricity in times with adequate wind and sun, and then feeding this power into the grid on windless and cloudy days.

Storage technologies are differentiated according to the amount of energy that can be stored and the length of time it can be stored. Battery storage systems (BSS) are examples of conventional methods that store energy for short-term periods (i.e., minutes or hours).

Pumped-storage power plants are employed, or hydrogen can be used as an energy vector when it comes to mass energy storage for more extended periods. BSS also help distributed generation operators to integrate and better utilize power

The total emissions profile of an LNG plant is now a key consideration for stakeholders. It can significantly impact whether or not a project reaches a final investment decision.

generation from renewable sources. They compensate for the volatility of renewable sources by either storing energy to be injected at a different time when demand or the price of energy is higher, or by quickly charging and discharging power to respond to surges and dips in renewable power availability, thereby ensuring a smooth output to the load. As a result, BSS contribute to optimizing energy costs, and their programmable and controllable fast response times improve matching supply and demand. The decision to incorporate energy storage is also crucial for sparing philosophies and redundancy to ensure high availability.

LNG operators typically design PGUs with a target availability of 99%+ because it is considered a utility that should always be available to the LNG train. To achieve this, the PGU is designed to a minimum $N + 1$ configuration, where N is the minimum number of gas turbines required for operation at 100%. In other words, PGUs are designed with a spare gas turbine that can either be running or offline, depending on the operating philosophy. A running spare will result in the gas turbines running at lower loads, away from the optimal load, for the highest efficiency. On the other hand, an offline gas turbine will require supplemental power by power augmentation or a battery when the gas turbine is started up. Otherwise, a load shedding procedure must be implemented to keep the LNG train operational.

Fuel system. In mechanically driven LNG plants, natural gas used during start-up and refrigerants that must be de-inventoried from equipment are sent to the facility's flare system for proper disposal. The contents are burned, and thermal energy is emitted into the atmosphere. However, these streams can instead be recovered and routed to the fuel gas system for conditioning. In such cases, the hydrocarbons will be utilized as a component of the blended fuel gas for the PGU.

Processing the streams through the fuel gas system is essential because of the heating value variation control inherently provided by the residence time designed in the fuel gas mixing drum. If the heating value rate of change varies outside of a prescribed range, then the gas turbine generators could be tripped. Consequently, the heating value of the recovered hydrocarbons must be known; so, too, will its impact on the overall fuel blend heating value to avoid exceeding the rate of change allowed for normal operation.

Moreover, a solution where recovery of these streams is envisioned requires additional kit and plot space not accounted for in a traditional LNG facility design. Thus, an integrated approach that includes an evaluation provides a more robust design and limits potential rework in the future.

Takeaway. Despite its growing role in the global energy mix, the carbon footprint of the LNG industry is being put under the microscope by activists, investors, lenders and regulatory bodies. The total emissions profile of an LNG plant is now a key consideration for stakeholders. It can significantly impact whether or not a project reaches a final investment decision (FID).

Electrification is an important pathway to the decarbonization of LNG facilities, providing immediate advantages in terms of higher-efficiency operations and lower emissions, while allowing for current or future implementation of renewables and green fuels (**FIG. 2**).

In the same way that refrigeration compressor mechanical drive gas turbines have been viewed as key to optimizing all aspects of LNG plant performance, the focus with eLNG is now shifting to the source of power generation, requiring a new, integrated approach to its design and specification. In light of this development, a strong case can be made to move toward a more holistic design approach where the required process and electrical systems are developed in parallel to enable total plant optimization. **GP**

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Case study: Optimization of LPG process at Ju'aymah NGL fractionation plant

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Optimization initiatives for the LPG Merox process were implemented to reduce operating cost and spent caustic generation at Saudi Aramco's Ju'aymah NGL fractionation plant. These initiatives included an increase in fresh caustic concentration and reuse of extractor spent caustic in the Merox process.

Also, a new catalyst was successfully used in the Merox system, with higher activity than a conventional catalyst. Benefits were derived from the new catalyst with regard to consumption rate, fresh caustic use and spent caustic generation.

The implementation of these initiatives contributed significantly to enhanced treatment efficiency, with a lower operating cost of $0.5 MM/yr, both from fresh catalyst and disposal of spent caustic. Spent caustic generation from the LPG Merox process was reduced by approximately 50%.

Process background. The Ju'aymah NGL fractionation plant has four identical fractionation trains. Each train consists of deethanizer, depropanizer and debutanizer columns; an amine treating facility to remove hydrogen sulfide (H_2S) and carbonyl sulfide (COS) from C_3 product received from the depropanizer overhead, and Merox processing units for mercaptan sulfur removal from C_3 and C_4 products, followed by the C_3 and C_4 dehydration system for moisture removal from the final product.

An opportunity was identified for the optimization of the LPG Merox process to reduce spent caustic regeneration, product sulfur and catalyst consumption, and therefore reduce plant operating cost. This article describes the initiatives implemented to optimize the Merox process and achieve these benefits.

Merox process overview. The Merox process consists of three sections, as shown in FIG. 1. The first section consists of prewash drums to remove H_2S and COS left over from the amine unit. Separate prewash drums are in place for both C_3 and C_4 products. Followed by prewash drums, the second section is the extractor, which contains C_3 and C_4 mercaptan extractors, and the last section is the combined Merox regeneration for the regeneration of mercaptan-rich caustic collected from both of the extractors.

The prewash drums are operated at a caustic concentration of 5 wt%–10 wt%, while the extractors are operated at a caustic concentration of 8 wt%–12 wt%. In extractors, mercaptan reacts with caustic and forms sodium mercaptide (Eq. 1), which is oxidized in the regeneration section for conversion back to caustic and disulfide oil (Eq. 2).

$$RSH + NaOH \rightarrow NaSR + H_2O \quad (1)$$

$$NaSR + \tfrac{1}{4}\, O_2 + \tfrac{1}{2}\, H_2O \rightarrow NaOH + \tfrac{1}{2}\, RSSR \quad (2)$$

The overall Merox process reaction can be written as shown in Eq. 3:

$$RSH + \tfrac{1}{4}\, O_2 \rightarrow \tfrac{1}{2}\, RSSR + \tfrac{1}{2}\, H_2O \quad (3)$$

In the Merox process, intermittent liquid Merox catalyst is injected into the rich caustic before the oxidizer. In the oxidizer, mercaptide is oxidized to disulfide oil and caustic, which are separated in the disulfide separator drum. In the overall Merox process, the caustic and Merox catalyst concentration is reduced with time due to the dilution effect.

To maintain optimum concentration, partial caustic replacement is carried out every couple of days. Due to a high system volume, this generates a large quantity of spent caustic from the Ju'aymah NGL facility every year. Shipment of the spent caustic is a costly measure and has a significant impact on plant net direct expenditure.

Process challenges and optimization opportunity. In addition to high spent caustic generation, other Merox process challenges included frequent system troubleshooting to control the product sulfur, high Merox catalyst consumption rate and high catalyst expense.

An opportunity was identified for the optimization of the Merox process to reduce product sulfur, spent caustic regen-

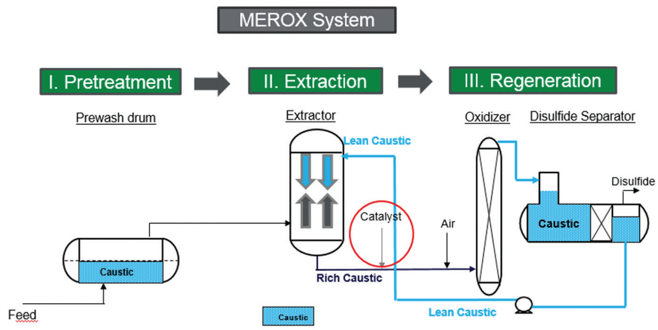

FIG. 1. Merox process overview.

eration and catalyst consumption and, therefore, reduce plant operating cost. The plant engineering team generated several in-house initiatives to optimize the process, several of which were successfully completed:

- New catalyst deployment in the Merox system
- Increase of fresh caustic concentration from 25 Baume (°Be) to 35°Be
- Reuse of extractor spent caustic in prewash drums
- Development of new training and awareness sessions for personnel
- Implementation of an online monitoring dashboard.

Each of these initiatives are described in detail in the following sections.

New catalyst deployment. To promote the oxidation reaction in the Merox regeneration section, liquid Merox catalyst is added intermittently. Catalyst is initially loaded after the unit shutdown for testing and inspection, with fresh caustic at an approximate concentration of 200 ppm. This catalyst is diluted with time after every caustic dumping; consequently, makeup catalyst is added to maintain the optimum concentration.

In 2018, plant personnel deployed a new catalyst to enhance the Merox system's overall performance. Initially, a pre-pilot test assessment was completed with Saudi Aramco Central Engineering Department's Process and Control Systems Department. The test ran the new catalyst in Modules 1 and 2, with fresh caustic, after testing and inspection. A pilot test was carried out for 6 months, and the Merox system operating parameters, including the product sulfur and catalyst consumption, were monitored.

The overall experience with the new catalyst was very good, and its activity was found to be significantly higher than the conventional catalyst, as confirmed by the catalyst consumption pattern. Due to this higher activity, the catalyst consumption rate was reduced significantly. Initial catalyst loading was approximately 50% lower compared to the old catalyst, and the cumulative catalyst consumption rate was reduced by more than 50% in normal operation (FIG. 2). Product sulfur was reduced by 6 ppm, compared with the old catalyst (FIG. 3).

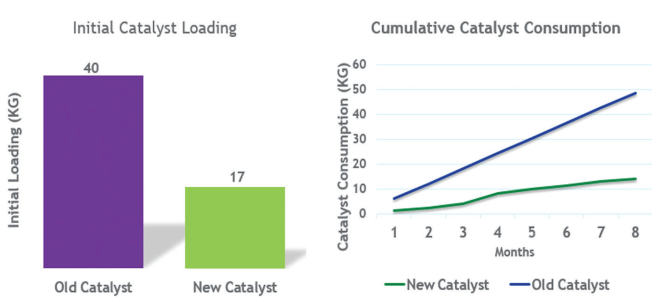

FIG. 2. Catalyst consumption rate.

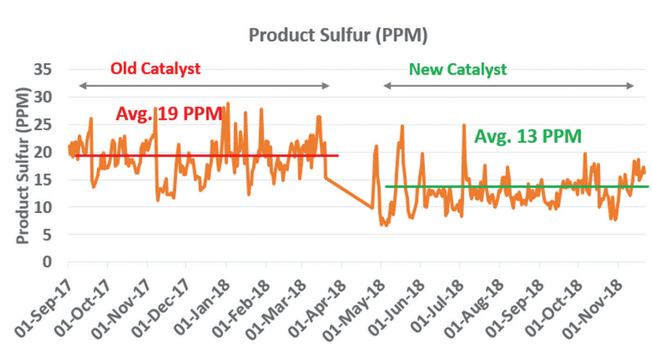

FIG. 3. Product sulfur before and after new catalyst use.

With the use of the new catalyst, spent caustic generation was reduced by approximately 20% due to the efficient regeneration of mercaptide in the Merox regeneration section; fresh caustic consumption was also reduced by the same rate. The subject catalyst per unit price was also significantly lower than the conventional catalyst. The use of the new catalyst significantly reduced the overall operating cost. Additionally, since the use of catalyst methodology and procedures were the same as for the old catalyst, no additional process modification or training for plant operators were required.

Increasing fresh caustic concentration. The next Merox process optimization initiative increased the fresh caustic concentration from 25°Be to 35°Be.

As described in the previous section, the caustic concentration comes down with time due to the dilution effect in the Merox process during normal operation; therefore, partial caustic is dumped from the system and refreshed with fresh caustic. From startup, the plant's normal practice was to import the fresh makeup caustic as a concentration of 25°Be, which is approximately 19 wt%.

Approximately 20% of the total caustic from the system with a 10 wt% concentration is replaced with fresh caustic at 19 wt% concentration; therefore, the caustic concentration was increased from 10 wt% to 11.8 wt%. In normal operation, this concentration would last for approximately 4 weeks before the next caustic dumping was required.

One process improvement was implemented to address the caustic concentration. By increasing the concentration from 25°Be to 35°Be, which is 19 wt% to 29 wt%, the concentration in the system was increased by approximately 1.8 wt%, compared with the 25°Be fresh caustic use. Increasing the caustic concentration in the system lasted longer than it had previously; therefore, the caustic dumping frequency was reduced by around 20%. The total caustic dumping volume was also reduced by 20% in 1 yr, which positively impacted plant operating cost.

Extractor spent caustic reuse in prewash drums. As explained in the Merox process overview section, the prewash drums operate with a caustic concentration limit of 5 wt%–10 wt%, with 30%–50% of liquid level to avoid spent caustic carryover into the extractor. The objective of this lower concentration of caustic in the pretreatment area is to remove the remaining H_2S and COS from the feed before entering the extractors.

The caustic in the prewash is not regenerated and is spent with time, where caustic molecules would be no longer free to react with the remaining H_2S in the feed. When the concentration reaches lower than 5 wt%, it is completely removed from the system and refreshed with fresh 10 wt% caustic.

Plant personnel realized that the extractor spent caustic, after being reduced to lower than 8 wt% concentration, was still suitable for prewash drum use. The plant operations team be-

gan practicing the transfer of spent caustic in the extractors to the prewash drums for full reuse. This initiative reduced spent caustic generation by a significant amount.

Additional training and online monitoring. The last initiative involved the use of awareness programs, additional training and the implementation of an online monitoring dashboard.

The plant engineering team realized the need for greater awareness among young engineers and operators, as well as more training to operate the Merox system efficiently. Caustic dumping and refreshing were considered preferred practices to control product sulfur; therefore, a learning environment was created to troubleshoot the Merox process and provide more in-house training and workshops to the operators and engineers.

An operator handbook, *Sulfur Troubleshooting Guide,* was published with simplified system troubleshooting charts. Additionally, an online troubleshooting dashboard in the plant information and distributed control systems was developed to be accessible to engineers and operators for critical parameters for online monitoring, as well as comparison with standard operating range parameters, for easy identification and correction of out-of-range parameters.

In addition to the Merox system, the amine system was included in this dashboard. Amine system parameter monitoring is equally important to avoid H_2S and COS carryover into the Merox system. This initiative significantly helped reduce product sulfur without frequent caustic dumping from the system.

Takeaway. The successful implementation of several initiatives to optimize the LPG Merox process contributed significantly to the enhancement of Merox treating efficiency by reducing operating cost by more than \$0.5 MM/yr, both from the use of fresh catalyst and the disposal of spent caustic.

Additionally, these initiatives reduced spent caustic generation from the LPG Merox process by approximately 50%. All of the projects were implemented without process modification or CAPEX by the company. **GP**

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The path to midstream profitability: Retrofit to improve returns

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The U.S. shale revolution has matured. Due to higher oil prices, shale production has shifted from natural gas basins to oil-rich basins. Oil-focused drilling typically produces associated gas as a byproduct, which is a blend of natural gas and natural gas liquids (NGLs). The U.S. is the world's largest NGL producer, supplying 11 MM metric t, or 29% of the global NGL supply, earlier this year. To manage these volumes, the U.S. has more than 400 gas processing plants that separate the associated gas into distinct streams of NGLs and natural gas.

Most gas processing plants recover NGLs, such as ethane and propane, using a process technology[a] developed in the late 1970s. Downstream of the gas processing plant, ethane and propane are feedstocks for producing ethylene and propylene. The abundance of these products has boosted U.S. petrochemical production and has driven the dramatic increase in petrochemicals exports to the global market.

Pivoting to profitability. Oil markets recently have faced three significant downturns: the 2007 global financial crisis, a global price war in 2014 and the severe drop in demand caused by the COVID-19 pandemic in 2020. Prior to COVID-19, particularly between 2018–2019, the growth rate of U.S. oil production had already begun to slow because of years of oil production growth without a focus on profitability. Overall, investor returns of Standard & Poor's 500 index dramatically surpassed that of the exploration and production sector, causing investors to pressure shale producers to focus more on slower but more profitable growth.

Gas processors responded by modifying existing assets to improve returns. In a capital constrained environment, gas processors can typically upgrade or retrofit their existing equipment for less than 5% of the total installation cost of a new gas processing plant. By doing so, they can efficiently improve product recovery, operational flexibility, energy efficiency and plant capacity. Given the historical volatility of the oil market, gas processors can ensure profitability by increasing the flexibility of their assets in preparation for a range of market conditions.

The value of ethane and propane. As the global economy recovers from COVID-19, the growth of U.S. oil production is forecast to be slower than it was pre-COVID-19, reducing the growth of NGL supply. Simultaneously, global demand for petrochemical feedstocks (e.g., ethane and propane) is forecast to surpass pre-COVID-19 growth rates.

Global ethylene demand is forecast to increase 5%/yr to approximately 200 MM metric t by 2025. Forty percent of ethylene production will be derived from ethane as a feedstock. Global ethane demand is forecast to increase to 5.2 MMbpd in 2025 (FIG. 1), a large portion of which is driven by U.S. demand centers along the U.S. Gulf Coast. To match the increase in ethane demand, U.S. gas processors will recover more ethane from associated gas. During this period, U.S. ethane supply will increase by 5%/yr, ultimately supplying 49% of global ethane in 2025.

Ethane will be more valuable for gas processors with close geographical proximity to the major demand centers—such as the U.S. Gulf Coast, where most U.S. petrochemical facilities and export terminals are located. For example, the Permian Basin—the fastest growing oil production basin in the U.S.—is near U.S. Gulf Coast export terminals and local demand centers. In addition, the Permian Basin has lower production costs due to an abundance of oil reserves and established infrastructure. Ethane pricing at the Mont Belvieu terminal will continue to increase through 2025. Because of these economic advantages, gas processors in the Permian Basin have shifted their operations to ethane recovery—the practice of recovering ethane and other high-value NGLs, such as propane.

Conversely, due to smaller profit margins, the same may not be true for capturing ethane in basins with higher production costs (e.g., the Niobrara Basin in northeast Colorado and southeast Wyoming). These basins tend to operate under ethane rejection, where ethane is not recovered but simply left in the natural gas stream to be sold at lower natural gas prices. For this reason, ethane's price floor is the price of natural gas. The profitability of either operation mode—ethane rejection or ethane recovery—will depend on the market price of ethane at a point in time. As the price of ethane continues to increase, there are new opportunities for swing basins to switch from ethane rejection to ethane recovery to capture ethane's additional value.

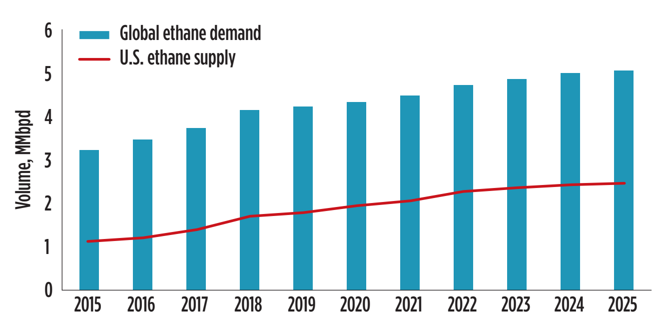

FIG. 1. Global ethane demand (MMbpd), 2015–2025. Source: IHS Markit.

Global propylene production via propane dehydrogenation, a process that utilizes propane as a feedstock, is projected to grow 14%/yr to 25 MM metric t in 2025. The increase in propylene production equates to a 5%/yr growth in propane demand, which will result in high propane prices for the foreseeable future in all U.S. regions. By 2025, global propane demand will be approximately 6.4 MMbpd (FIG. 2), 37% of which will be supplied by the U.S. Therefore, efficiently capturing as much high value propane as possible is critical for improved profitability for U.S. gas processors in all basins.

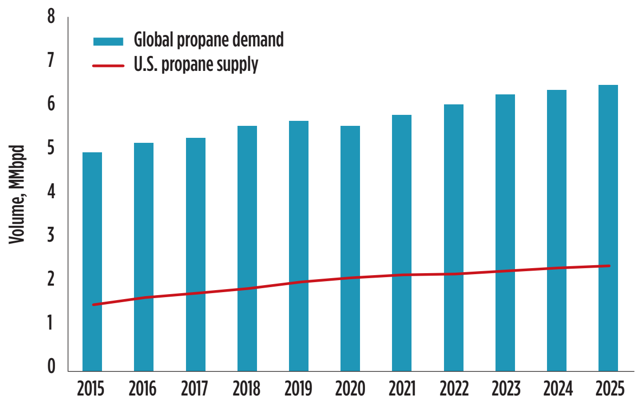

FIG. 2. Global propane demand (MMbpd), 2015–2025. Source: IHS Markit.

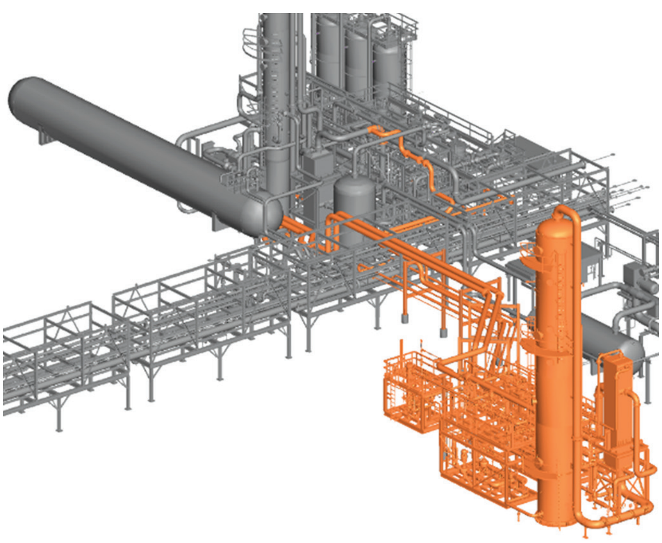

FIG. 3. An example of a retrofit modification with new equipment highlighted in orange.

FIG. 4. Brazos Midstream's 200-MMft³d gas processing facility. Source: Brazos Midstream.

Optimization through retrofits. Gas processors have started retrofitting existing assets to further optimize operations. Retrofitting a gas processing plant upgrades the existing design to more advanced technology, which improves ethane and propane recovery, operational flexibility, and increased plant capacity. A retrofit can improve ethane rejection, ethane recovery or both. In addition, a retrofit can increase the plant's capacity and energy efficiency. Given the unpredictability of the oil industry, a retrofit is a proven, high return method to increased profitability.

The best retrofit technology depends on the gas processor's feedstock, contracts, location and regional market conditions. For example, the authors' retrofit technologies can increase NGL recovery, reduce energy consumption, improve operating flexibility and increase processing capacity. Typically, a retrofit can recover nearly 100% propane and 99% percent ethane when operating in ethane recovery. When in full ethane rejection, more than 98% ethane can be left in the natural gas stream, while recovering 99%, or more, propane. Furthermore, the plant's capacity can often be increased above 120% of the original plant design, depending on existing equipment limitations. The required compression horsepower of the retrofit will vary but, due to the focus on energy efficiency, retrofits can fit within the existing compression limits of the plant.

Many of the authors' retrofit technologies are modular designs that provide easy connection to existing gas processing plants, minimizing the changes to the existing equipment and on-site construction required during the installation. The equipment that may be installed during a retrofit is highlighted in FIG. 3. The efficiency of a modular design can reduce plant downtime to less than 1 wk and allows the facility to recoup the installation cost in a shorter period. Depending on the gas processor's contracts and market conditions, some retrofits can be paid off in less than 1 yr.

In the Permian Basin, Brazos Midstream is retrofitting two 200-MMft³d gas processing plants, each using the authors' company's proprietary gas processing technology[a]. These ret-

TABLE 1. Retrofit study results for a 300-MMft³d gas processing plant

Recovery	Initial design	Lower CAPEX retrofit	Higher CAPEX retrofit
Ethane, mol%	71	90	97
Propane, mol%	98	100	100
Incremental ethane production, bpd	Base	8,000	10,900

TABLE 2. Results of a retrofit project on an 80-MMft³d gas processing plant

	Ethane recovery		Ethane rejection	
Recovery	Initial design	Retrofit	Initial design	Retrofit
Ethane, mol%	72	95	N/A	1
Propane, mol%	98	100	N/A	100
Power requirements, hp	6,100	4,100	N/A	4,200

rofits provide greater flexibility with improved product recovery when operating in ethane recovery or rejection modes. The standard process technology[a] design can switch from ethane recovery to rejection but, with the retrofit, the recovery of ethane and propane will be nearly 100%. The plants also can increase production volume as market conditions dictate. The retrofits are scheduled to be fully operational by the end of this year.

Retrofit studies. Globally, the authors' company has retrofitted more than 50 gas processing plants. The ideal retrofit to maximize value add will be customer specific. A recent study was completed on a 300-MMft^{3}d gas processing unit. The client's focus is to increase ethane recovery in anticipation of future ethane demand, while minimizing capital expenditure (CAPEX). The study determined that a low CAPEX retrofit would improve ethane recovery from 71% to 90%. Installing additional equipment would further increase ethane recovery to 97% (TABLE 1). For this facility, improving ethane recovery to 90% or 97% would increase annual revenue by an estimated $24 MM or $34 MM, respectively. These values do not include the additional revenue that would be generated from the improved propane recovery. It is projected that both the lower and higher cost retrofits would be paid off in less than 1 yr.

A retrofit was completed on an 80-MMft^{3}d unit that was originally designed to recover 72% ethane and 98% propane. This facility was not capable of switching to ethane rejection operation. The authors' company retrofitted the unit, while satisfying the customer's four primary objectives:

1. Enable flexibility to operate efficiently in either ethane recovery or ethane rejection modes
2. Increase propane recovery to above 99% in all operating modes
3. Maximize the use of existing equipment to reduce CAPEX
4. Minimize plant downtime for installation.

The authors' company implemented a retrofit that provided the customer with a stable and simple operation that exceeded design expectations. As detailed in TABLE 2, ethane recovery was increased from 72% to 95%, while reducing power consumption by 33%, which reduced the plant's operating expense and significantly improved profitability. The facility is now capable of running effectively in either ethane rejection or ethane recovery modes. To minimize downtime, new equipment was delivered as a bolt-on package that was easily integrated into the existing unit.

By combining the gas processing technology with modular equipment design, the authors' company is uniquely positioned to provide customized retrofits in a cost-efficient manner. As global petrochemical demand increases, ethane and propane prices are forecast to rise. Gas processors that have upgraded their plant technology will be in a better position to maintain strong profitability. **GP**

NOTES

[a] Gas Subcooled Process developed by Ortloff Engineers. Ortloff was acquired by Honeywell UOP in 2018.

Simulation and optimization study for a condensate recovery system

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This article discusses a simulation and optimization study for a low-pressure condensate recovery system used on an oil and gas platform for the extraction of valuable hydrocarbons from flare gas for sale. Due to the possibility of hydrates formation, methanol is used as a hydrate inhibitor in this case.

A base case simulation model was constructed, using commercial software[a] to identify high-risk points where methanol injection is to be performed. The use of methanol entails increased operating cost; therefore, it is necessary to determine the relationship between the methanol flowrate and condensate recovery, as well as the operating temperature of the cold separator. The latter has a significant effect on the condensate recovered.

The optimum methanol flowrate was determined by maximizing the profitability of the system through an optimization model, which is a linear program (LP). The LP model was solved using Microsoft Excel Solver. The optimum temperature was determined to be –21.6°C, with a net profit of $18,000/d. The system was determined to have an emissions avoidance of approximately 39,000 tpy of CO_2 through the recovered condensate.

The three major sources of natural gas in the oil and gas industries are oil, gas and condensate wells. Natural gas found in oil wells is termed as associated gas and is either dissolved in the oil phase or remains as cap gas above the oil.[1,2] The associated gas may have a similar composition to that of natural gas, but the specific composition can vary significantly based on the production location and the properties of the well.

Heavy investment has been reported for the further processing of this gas. However, some of this gas is flared due to technical or economic limitations. It has been reported that more than 17,000 oil production facilities worldwide have flared approximately 140 Bm^3/yr of natural gas, resulting in the emissions of more than 350 MMt of CO_2 (along with other pollutants).[2]

The oil and gas industries continue to work diligently toward CO_2 abatement. The contribution from large industry players such as ExxonMobil, Shell and BP contributed to a CO_2 reduction of about 12% between 2010 and 2015.[3] The World Bank has initiated "Zero routine flaring by 2030" and encourages global oil companies to endorse the efforts in mitigating CO_2 emissions.[4] Different attempts have been reported to reduce flaring, e.g., re-injection into oil reservoirs to maintain pressure and to increase oil recovery,[2] zero flaring during startup,[5] condensate recovery system,[6] etc. In particular, the low-pressure condensate recovery system (LP-CRS) has been reported to extract valuable hydrocarbons from flare gas under low pressure with minimal rotating equipment.

Background. The Tembikai oil and gas field is located 150 km offshore the east coast of Peninsular Malaysia. It includes three production wells and a central processing platform (CPP). The oil from the CPP is sent to a floating storage and offloading (FSO) vessel through a flexible subsea pipeline.[7] The associated gas from the Tembikai CPP has been flared in the traditional practice. The flared gas consists predominantly light hydrocarbons [e.g., methane (C_1) to butane (C_4)] and significant quantities of pentane (C_5) and heavier components (C_5+).

An LP-CRS has been installed that is specifically tailored for the extraction of valuable C_5+ components. The LP-CRS makes use of a turboexpander coupled with a Joule Thomson (JT) valve, allowing the system to perform at a lower temperature to maximize condensate recovery.[8] In this work, optimization is performed for the LP-CRS to minimize the amount of methanol required for the prevention of hydrates formation in the pipelines.

The associated gas is produced from a reservoir where it stays in equilibrium with water—the low operating temperature may lead to the formation of hydrates, which are formed through the crystallization process where the gas molecules are trapped in crystalline cells formed from the hydrogen bonds of water molecules. A high-pressure and low-temperature environment assists the formation of hydrates.[9] Hydrates reduce the diameter and ultimately block the pipeline, downhole tubing, tree and manifold piping, flowlines and risers. The plugs can be difficult to locate and remove, leading to significant losses in production and revenues.

Dehydration can be used to eliminate the formation of a condensed water phase.[10] However, in certain cases, dehydration may be impractical or economically feasible. Therefore, inhibition is commonly used to prevent the formation of hydrates. Thermodynamic inhibitors—typically methanol or glycol injection—lower the hydrate formation temperature at a given pressure. For cases of below ambient temperature, methanol injection is preferable to avoid the viscosity and freezing issues of glycol. However, the methanol content should be kept at its minimum level, as it incurs additional operating cost to the system. It also impacts downstream processes and presents environmental limitations on overboard discharge.[11]

Base case process simulation. FIG. 1 shows the base case simulation model of a condensate recovery system simulated using commercial software[a] using Peng-Robinson as its thermodynamic model. Peng-Robinson presented the worst-case

scenario, while other models displayed ice formation only under similar process conditions at the respective high-risk points. The inlet stream consisted of associated gas with a volume of 8.0 MMft³d, with a temperature and pressure of 81.6°C and 6 bar, respectively. The composition of the gas stream is given in TABLE 1, while the specifications of the units are given in TABLE 2.

The associated gas entered the compressor side of the turboexpander and left at 9 barg. The compressed gas then entered an air cooler, where it was cooled to 45°C. The outlet stream was then fed to the hot side of Plate Heat Exchanger 1 (HE-01), where it was cooled further to 10.5°C before passing through an inline separator. The latter removed 98% of the water present in the feed at its bottom stream (Sep. Bottom), along with some low-quantity heavy end hydrocarbons (C_4s and C_5+). This stream was discharged from the system.

Next, the top product from the separator was passed through the hot side of plate HE-02, where it was cooled to –21.6°C. This chilled stream was then sent to the cold separator, where the heavy end hydrocarbons were recovered as the main product of the system at its bottom stream. The top product of the cold separator was passed into the expansion section of the turboexpander, where its pressure was reduced to 1.7 barg.

The energy recovered from the expander was supplied to the compressor section of the turboexpander for the compression of feed. The compressed stream was then passed on to Plate HE-02 to cool the hot streams. This stream was then combined with the bottom stream of the suction scrubber (Scrub. Bottom), which consists mainly of water. Finally, this stream was used to cool the hot streams in Plate HE-01 before it was sent to flare.

TABLE 1. Composition of inlet stream

Component	Mole fraction
Nitrogen	0.01
Carbon dioxide	0.03
Methane	0.64
Ethane	0.08
Propane	0.09
i-Butane	0.03
n-butane	0.02
i-Pentane	0.01
n-Pentane	0.01
n-Hexane	0.01
Water	0.07

TABLE 2. Specifications for process units

Equipment	Section	Specified temperature, °C	Pressure drop, barg
Turboexpander	Compressor inlet	81.61	–
	Compressor outlet	–	
	Expander inlet	–	
	Expander outlet	–	
Air cooler	Inlet	–	0.2
	Outlet	45	
Plant HE-01	Hot side inlet	–	0.2
	Hot side outlet	10.46	
	Cold side inlet	–	0.2
	Cold side outlet	–	
In-line separator	–	–	–
Plant HE-02	Hot side inlet	–	0.2
	Hot side outlet	–	
	Cold side inlet	–	0.2
	Cold side outlet	–	

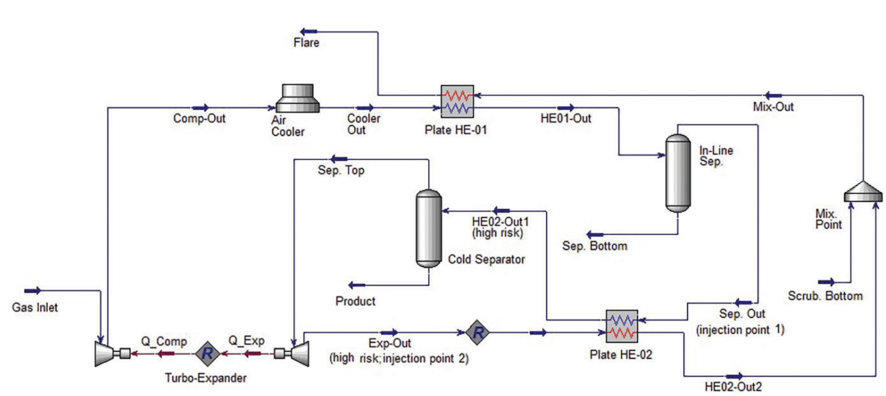

FIG. 1. Simulation flowsheet of condensate recovery system.

Methanol injection. Before determining the methanol intake, the high-risk points in the system were first identified. These are locations where hydrate formations were likely to occur. First, the outlet stream of Plant HE-02 (i.e., HE02-Out1 stream) was found to exhibit Type II hydrate formation, according to stream analysis results of the simulation model. Note: Type II hydrates will form in the case of gas mixture,[10] and they are relatively more stable and form at a higher temperature compared to Type I. Since stream HE02-Out1 is the inlet stream to the cold separator, it was assumed to have the same condition as the separator outlet streams (i.e., the Sep. Top and Product stream).

Therefore, methanol injected at the upstream of HE02-Out1 is sufficient to suppress hydrates formation at the separator inlet and outlet streams. Another high-risk point was found at the expander outlet stream (i.e., Exp. Out) that also exhibits Type II hydrate formation.

The next step was determining the location for methanol injection points. The first high-risk point (i.e., HE02-Out1) is the outlet stream of Plate HE-02. Instead, the inlet stream of the latter (i.e., Sep. Out) was chosen as the first injection point. Even though the temperature of the Sep. Out stream was not low enough to form hydrates, hydrates will form as soon as the temperature is lowered to –21.6°C in Plate HE-02.

Apart from preventing hydrates formation in the HE01-Out1 stream, the tubes within Plate HE-02 are also being protected. Conversely, as the stream temperature continues to drop from –21.6°C (Sep. Top) to –57.0°C (Exp. Out, due to

the expansion), the Exp. Out stream was suggested as the second methanol injection point. A concern was that the first methanol inject point located upstream (Sep. Out stream) may not be sufficient to suppress the formation of hydrates at this stream, due to its lower temperature. It is important to ensure good operating condition of Plate HE-02, as it will affect its outlet stream (i.e., HE02-Out2 stream). This then affects the cooling of the HE01-Out stream, and subsequently the Sep. Out stream, in turn impacting the recovery of hydrocarbons in the main product stream.

The base case simulation model was then used to determine the required amount of methanol for both injection points. Analysis showed that 0.35 tpd of methanol is required for the Sep. Out stream (Injection Point 1) and 0.39 tpd for the Exp. Out stream (Injection Point 2). These flowrates should be optimized by considering economic performance.

Prior to the construction of the optimization model, important parameters that affect the systems should be determined. The simulation model revealed that the operating temperature of the cold separator had a significant effect on the methanol requirement and the condensate recovered, while the operating pressure effect was not significant. Therefore, the effect of operating temperature was examined though the simulation model, shown in FIGS. 2 and 3.

FIG. 2 shows the relationship of the methanol flowrate vs. the operating temperature for the temperature range of −21.6°C to −1.6°C (with temperature intervals of 5°C). The amount of methanol required at the first injection point decreased by only 0.07 tpd when the temperature increases to −1.6°C. However, the methanol required at the second injection increased drastically from 0.39 tpd to 1.19 tpd for the same temperature rise. When the temperature increases, less water is condensed to the bottom of the cold separator (product stream); instead, more water is vaporized to the Sep. Top stream and enters the expander side of the turboexpander. More water content is found in the Exp. Out stream, which increases the tendency of hydrates formation in this stream, leading to an increased requirement of methanol in this stream.

In FIG. 3, the amount of condensate recovered against the temperature of the cold separator is determined. The HE02-Out1 stream is comprised of approximately 40% of hydrocarbons heavier than C_2. Therefore, the operating temperature of the cold separator greatly affects the amount of its recovered condensate in the main product stream. It was observed that a higher amount of condensate was recovered at a lower operating temperature of the cold separator. The amount of condensate recovered was observed to drop by 80 bpd for every 5°C decrease in temperature (FIG. 3).

Optimization model. The values obtained from the simulation model were incorporated into a linear regression (FIGS. 2 and 3), and their equations were used as optimization constraints. Eq. 1 shows the methanol required against temperature (first injection point):

$$F_{M1} = -0.003T + 0.2857 \qquad (1)$$

Eq. 2 shows the methanol required against temperature (second injection point):

$$F_{M2} = 0.0394T + 1.1985 \qquad (2)$$

Eq. 3 shows the condensate recovered against temperature:

$$F_{Cond} = -12.984T + 85.846 \qquad (3)$$

where F_{M1} (tpd) and F_{M2} (tpd) represent the methanol flowrate at injection Points 1 and 2 with respect to temperature; F_{Cond} (bpd) is the condensate recovered with respect to T (temperature in °C), which represents the operating temperature of the cold separator.

The optimization objective is given by Eq. 4, which is essentially the difference between condensate sales and the methanol cost:

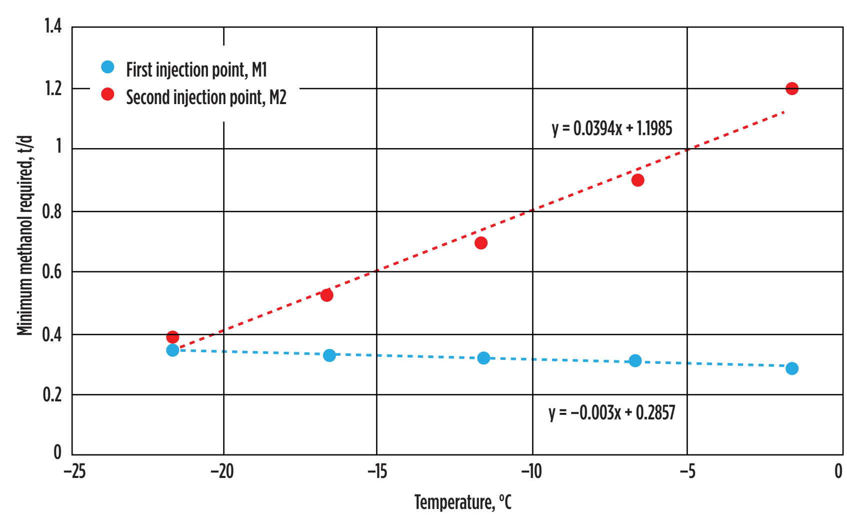

FIG. 2. Graph of methanol flowrate against temperature.

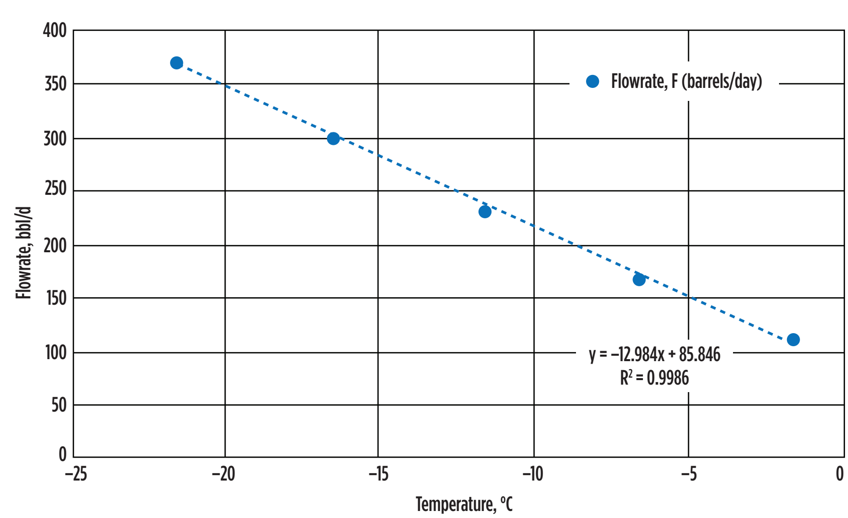

FIG. 3. Condensate recovered at the cold separator against temperature.

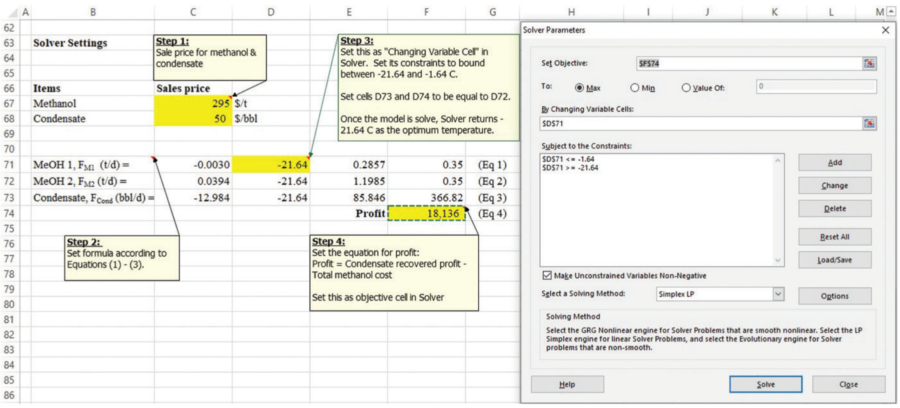

FIG. 4. Excel Solver for solving the LP model.

$$maximize\ P = (F_{Cond} \times C_{Cond}) - ((F_{M1} + F_{M2}) \times C_M) \quad (4)$$

where P (\$/d) represents the profit generated per day; C_M (\$/t) and C_{Cond} (\$/t) represent the unit cost of methanol and unit price of condensate, respectively. C_M was taken as \$295/t and C_{Cond} was given the value of \$50/bbl.[12,13] Note that the above model is a linear program (LP), which may be solved easily with any commercial solver, including Excel.

The objective in Eq. 4 was solved, subjected to the constraints in Eqs. 1, 2 and 3, yielding the maximum profit of \$18,136/d (see snapshot in **FIG. 4**). The model determined that the profit corresponded to 366.8 bpd of condensate recovery (**FIG. 4**). Verification with the simulation model determines that the condensate recovery is actually 367.8 bpd. Note that for this case, the amount of condensate recovered is the dominant factor of the profit, since the price of condensate is much higher than that of methanol—in terms of flowrate, the flowrate of condensate is again higher than that of methanol. Therefore, the profit obtained is solely dependent on the condensate flowrate.

However, even if the condensate flowrate is ignored, **FIG. 2** shows that a lower cold separator inlet temperature lowers the amount of methanol required. This scenario is justifiable with the increased amount of water and hydrocarbons (C_3, C_4) being condensed from the cold separator as the temperature decreases. Therefore, the amount of methanol required in the Exp. Out stream drops significantly as the temperature of the HE02-Out1 stream decreases.

Based on the results from the optimizaion model, it was concluded that the optimum operating temperature for the cold separator should be –21.6°C because this temperature recovers the highest amount of condensate with the minimum amount of methanol consumed. It is also worth noting that the condensate recovered by the LP-CRS corresponded to a significant reduction of 39,000 tpy of CO_2. This was determined by performing a complete combustion of the recovered condensate (367.8 bpd) using a conversion reactor model in the commercial software[a] (assuming 365 d/yr).

Takeaway. The study provided an optimization study for a condensate recovery process with an LP-CRS. High risk points and their respective methanol flowrates required for inhibition were determined. The report also provided the study of methanol flowrate and condensate recovered as the function of temperature. The mathematical model allows the user to maximize profits gained from condensate recovery, while minimizing the methanol flowrate. The optimized temperature of –21.6°C was obtained with the aid of Excel Solver. The maximum profit gained was calculated as \$18,136/d, and it was determined that at least 39,000 tpy of CO_2 can be prevented through this system. **GP**

NOTES

[a] AspenTech's HYSYS V10

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Free water prediction for corrosion protection of NGL pipelines

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Natural gas liquids (NGL) separated from natural gas carry varying concentrations of water in the form of free water, water emulsion and soluble water. Typically, free water will form once the total water concentration of NGL rises above its solubility level, and it can be separated out more efficiently by separators or coalescers. The solubility of water in the NGL is dependent on the fluid temperature and on the composition of the condensate; as a consequence, it typically ranges between 50 ppmw and 650 ppmw. Moist NGL becomes cool, and the dissolved water in it can freeze and eventually block the pipeline. Free water can also combine with corrosive gases like H_2S and CO_2 to form acids, which will often lead to corrosion issues, shortened pipeline life and significant fouling of condensate handling and downstream facilities.

Practical measurement of the solubility of water in NGL at varying process temperatures is difficult due to the highly volatile nature of the liquid and the labor-intensive nature of the process. However, this information is vital for the design and troubleshooting of gas processing facilities, for the control of free water formation in pipelines, and for corrosion protection at gas processing plants.

The study presented in this article shows how the solubility of water in NGL at varying process temperatures can be calculated based on the NGL composition and the water saturation values of the pure components present within it. The obtained moisture results were validated with the standard Karl Fisher moisture content determination method. The developed solubility curve can guide plant engineers in understanding the importance of improving the condensate dehydration step and the selection of appropriate technology or technologies upfront, so that they can remove free water early in the process.

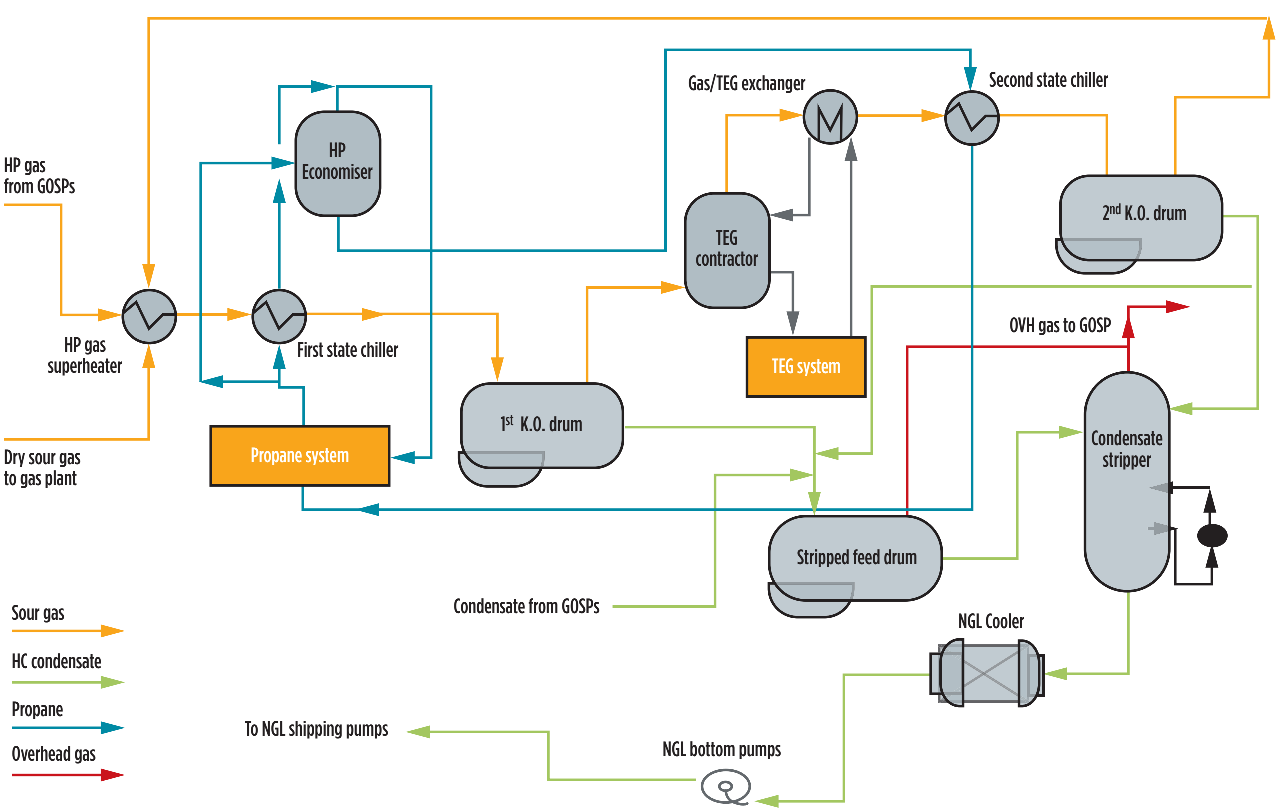

FIG. 1. Typical NGL process flow diagram in central processing facility.

Introduction. NGL, also known as hydrocarbon condensate, are light hydrocarbons in the same family of molecules as natural gas and crude oil that have condensed from the gaseous state of natural gas into a liquid state. This condensation may occur naturally in the wellsite when pressure is reduced, or at the surface. Aboveground, the presence of NGL in natural gas (also called rich gas) is unstable, as heavier components will condense, while lighter components normally remain gaseous and will need to be separated in a gas processing plant, either by distillation or a refrigeration/cryogenic process.

In general, NGL is composed of hydrocarbons, such as propane, butane, pentane, hexane, etc., and compounds with more carbon atoms (such as pentane, or blends of butane, pentane and other hydrocarbons with additional carbon atoms) will exist as liquids at ambient temperatures.[1] Hydrocarbon condensate that does not require specific processing can be directly sent to the export pipeline; this is typically the case at offshore production platforms.

Destabilized NGL contain several impurities, such as free, emulsified and dissolved water; salts; acidic components; corrosion inhibitors; hydrate inhibitors (monoethylene glycol, methanol and kinetic hydrate inhibitors); solid particles (e.g., corrosion products, sand); and solid-like particles (e.g., waxes, gels).[2–5]

Various separation technologies are available to separate gas, impurities, water and salt content from un-stabilized condensate, such as gravity settlers, knockout vessels with mesh pads, electrostatic desalters and cartridge coalescers, cooling (refrigeration), oil absorption, adsorption (such as by activated carbon or silica gel) or membrane processes.[6–8] All technologies have specific features that make them suitable for a given set of operating conditions.

Also, integrated NGL recovery can employ different schemes, such as condensation by refrigerant or expander technology.[9] Usually, the raw natural gas feedstock from a gas well or a group of wells (oil wells, gas wells and condensate wells) or a gas oil separation plant (GOSP) is cooled to lower the gas temperature to below its hydrocarbon dewpoint at the feedstock pressure. This condenses a large part of the gas condensate hydrocarbons. The feedstock mixture of gas, liquid condensate, and water is then routed to a high-pressure separator vessel (three-phase separator) and coalescers, where the free water and the raw natural gas are separated and removed. The condensate is then treated by a stabilization process (condensate stripper column) prior to export.

A typical NGL process flow diagram is shown in **FIG. 1.** The stripper column enables the separation of the H_2S, moisture and light ends to stabilize and sweeten the NGL product. The condensate stripper reboilers provide the heat required to strip the H_2S, residual water and light ends out of the hydrocarbon condensate. This operation aims to reduce the vapor pressure of NGL by eliminating the light fractions to make it safe for storage at atmospheric conditions and for transportation.[5–8] The stabilized NGL is marketed under various forms, depending on the production rate, composition and available downstream markets and transportation network.

Water as a contaminant in hydrocarbon condensate has been classified into different categories—usually free, emulsified, dis-

TABLE 1. NGL composition, vol%

Sample point	C_1	C_2	C_3	IC_4	NC_4	IC_5	NC_5	C_6	C_7	C_8
GPP-1	1.28	60.04	24.45	3.75	6.84	2.14	1.24	0.17	0.09	0
GPP-2	0	0	44.13	7.1	27.2	6	9.2	6.30	0.07	0
GPP-3	0	0	35.04	7.69	25.1	6.64	9.38	16.11	0.04	0
GPP-4	1.28	60.04	24.45	3.75	6.84	1.62	1.33	0.57	0.08	0.04
GPP-5	0.5	47	33.3	3.4	10.7	1.3	3.1	0.49	0.14	0.07
GPP-6	0	0	49.11	7.05	16.65	5.16	6.82	15	0.18	0.03

TABLE 2. Water saturation values (ppmw) of pure components at varying temperatures

	Water saturation values of pure components at different temperature					
Chemical	**0°C**	**10°C**	**20°C**	**30°C**	**40°C**	**50°C**
Methane	6	16	40	81	137	210
Ethane	31	65	130	250	400	600
Propane	56	93	154	240	385	580
Isobutane	25	42	71	120	202	340
n-Butane	20	34	60	104	185	305
isopentane	33	55	92	155	260	436
n-Pentane	29	49	81	135	226	376
Hexane	32	57	101	179	317	561
Heptane	27	54	96	172	308	480
Octane	27	51	95	168	297	525
Nonane	25	40	63	102	164	262
Decane	24	37	57	88	136	231

solved and total water. Free water is not dissolved in NGL. In some cases, it is characterized as the bulk fraction of water that separates out more easily by separators or by using coalescers. Emulsified water is contained in small drops (typically 0.1 micron–50 microns) and is usually more difficult to separate. Soluble water is dissolved at the molecular level in the hydrocarbon phase.

Total water content in condensate downstream of the inlet separators is typically present in concentrations varying from a few hundred ppmw up to 5%. The salinity of the water contamination is determined by the formation of the water, and also varies significantly from a few hundred ppm up to a few hundred thousand ppm.[12] Quality specifications of the dehydrated condensate prior to the stabilizer are project-dependent and typically show free water concentrations ranging from 0 ppmv to < 100 ppmv.

Usually, free water is removed from the condensate by using coalescer technology, or by cooling and passing it through a high-pressure separator. Complete free water removal from condensate is often difficult due to the formation of stable condensate/water emulsions, caused by the presence of surfactants, such as hydrate inhibitors and corrosion inhibitors that lower the interfacial tension.[3] Also, the condensate contains water in the soluble form. The solubility of water in the condensate is dependent on the fluid temperature and on the composition of the condensate; as a consequence, it can range between 50 ppmw and 650 ppmw, typically.

When moist NGL becomes cool, it can freeze and block NGL lines or pipes. Moreover, the presence of dissolved water in NGL beyond its solubility level is not advisable for transport on pipelines, because any change in ambient temperature or line pressure can lead to the development of free water from dissolved water. The free water can easily combine with acid gases, which can lead to corrosion issues in the condensate storage tank and in the export pipeline. Corrosion of the export pipeline may also represent a major integrity issue that could lead to premature replacement of some sections of the pipeline, if left unattended. An accurate prediction of free water formation is essential to prevent such problems.

The solubility of water in hydrocarbons, even at ambient temperatures, can have great practical importance.[14] Based on empirical and theoretical considerations, formulations are available in the literature for the prediction of water solubility in hydrocarbons and its dependence on temperature.[14–21] The experimental method is the more accurate way of determining water content at varying process temperatures. In this study, solubility of water in NGL at varying process temperatures was measured by using the NGL composition and the saturation vapor pressure of the pure components present within it. The obtained values were confirmed with Karl Fischer titration. The resulting solubility curve can be used for predicting free water formation at NGL handling facilities.

Outline of experimental procedure.

The NGL used in this study were taken from the export pipelines of five different gas handling and processing facilities in Saudi Arabia.

NGL composition. NGL composition was measured using a gas chromatograph equipped with a thermal conductivity detector (TCD), as well as an open tabular column of 100 m × 0.25 mm inside diameter, fused silica-coated with 0.5-micron-bonded methyl silicon. A helium carrier gas linear flow was established at 30 ml/min. Injector and TCD detector temperatures were maintained at 220°C and 250°C, respectively. The column oven temperature was programmed for an initial temperature of 70°C held for 4 min, a ramp-up to 8°C/min, and a final temperature of 200°C for 16 min. The injection volume was 1 µL.

Total water content. The water content of samples from each location was subjected to Karl Fischer coulometric titration (KFT) by following ASTM E1064-16.[13] With this method, dissolved and free water can be measured together. The condensate sample was collected in a dry, high-pressure, coated stainless-steel bomb. In the laboratory, the initial weight of the sample was recorded, and the condensate sample was scrubbed through dry ethylene glycol (25 ml in a separatory funnel).

At first, the high-pressure gas associated with the condensate was allowed to bleed slowly through the solvent. When the pressure had dropped considerably, the bomb was inverted and the remaining liquid condensate percolated through the glycol. When the bomb was empty, a gas line was attached to the upstream valve, and dry argon gas was purged through the bomb while heating the bomb walls with an electrical hot air gun. This procedure ensured that residual moisture on the internal walls of the bomb was transferred to the glycol solution (**FIG. 2**). Two distinct solvent layers should form: a lower glycol layer and an upper condensable hydrocarbon layer. The glycol layer is run off for Karl Fischer (KF) coulometric analysis.

Determination of water saturation constant. To confirm the calculated water saturation constant (Cs), the result was checked with KF results. For this purpose, a sample from GPP-3 was used to prepare a water-saturated hydrocarbon sample (excess water observable) mixed continuously at 300 rpm with magnetic stirrer. Then, the sample and stirrer were placed in a chamber at a set temperature and given enough time to reach equilibrium (18 hr). A coulometric KF titration cell was prepared and set for conditioning. The accuracy of KFT was verified by checking the solution injections. The mass of 1 ml volume at set temperature (23°C) was measured and transferred to the KF chamber. **TABLE 5** shows the measured saturated moisture concentration by KFT.

Results.

Samples from six gas handing facilities were subjected to gas chromatography analysis. The composition (in vol%)

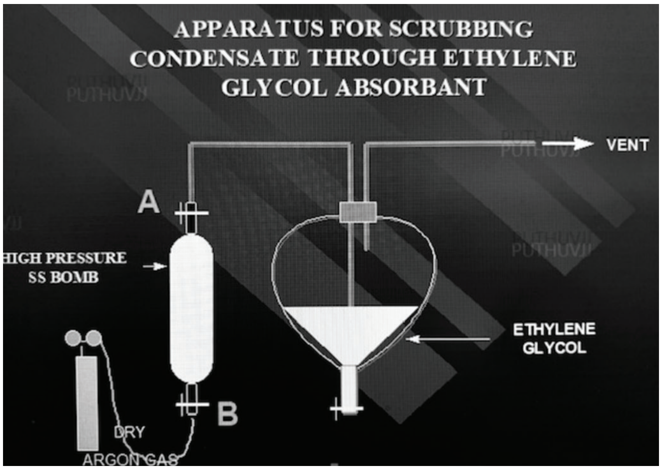

FIG. 2. Setup for NGL sample preparation for KFT.

TABLE 3. Water saturation value calculation for GPP-1

Component	Volume, %	Volume fraction	Molecular weight, g/mol	Volume fraction, mol wt	Weight fraction	Average Cs value					
						0°C	10°C	20°C	30°C	40°C	50°C
Methane	1.28	0.01	16.04	0.21	0.005	0	0.1	0.2	0.4	0.7	1.1
Ethane	60.04	0.6	30.07	18.05	0.476	14.8	30.9	61.9	119	190.4	285.6
propane	24.45	0.24	44.097	10.78	0.284	15.9	26.4	43.8	68.2	109.4	164.9
isobutane	3.75	0.04	58.124	2.18	0.057	1.4	2.4	4.1	6.9	11.6	19.5
n-butane	6.84	0.07	58.124	3.98	0.105	2.1	3.6	6.3	10.9	19.4	32.
iso-pentane	2.14	0.02	72.151	1.54	0.041	1.3	2.2	3.7	6.3	10.6	17.7
n-pentane	1.24	0.01	72.151	0.89	0.024	0.7	1.2	1.9	3.2	5.3	8.9
Hexane	0.17	0	86.178	0.15	0.004	0.1	0.2	0.4	0.7	1.2	2.2
Heptane	0.09	0	100.205	0.09	0.002	0.1	0.1	0.2	0.4	0.7	1.1
Octane	0	0	114.232	0	0	0	0	0	0	0	0
Nonane	0	0	128.259	0	0	0	0	0	0	0	0
Decane	0	0	142.286	0	0	0	0	0	0	0	0
Total	100	1		37.87	1	36.4	67	122.3	215.6	348.7	531.9

TABLE 4. Water saturation value (ppmw) of NGL samples at various locations

	0°C	10°C	20°C	30°C	40°C	50°C
Temperature	Water saturation values Cs, ppmw					
GPP-1	36.39	67.05	122.28	215.62	348.69	531.85
GPP-2	38.59	64.67	108.66	175.72	290.93	459.15
GPP-3	36.63	61.78	104.46	171.22	286.4	460.53
GPP-4	35.79	67.74	126.13	226.76	365.3	553.61
GPP-5	37.72	69.18	125.62	220.11	355.49	540.46
GPP-6	41.07	69.02	116.05	187.7	310.50	491.88

TABLE 5. Water saturation value by KFT

Sample point	Sample temperature, °C	Water content, ppmw
GPP-3, Sample-1	22	117.4 ± 0.8
GPP-3, Sample-2	25	137.7 ± 1.3
GPP-3, Sample-3	28	158.2 ± 1.5
GPP-3, Sample-4	32	194.8 ± 1.8

are given in **TABLE 1.** Énglin *et al.*[21] have tabulated water solubilities in wt% for a number of hydrocarbons at 10°C intervals between 0°C and 50°C. For this study, we selected only water saturation values of pure components (i.e., methane, ethane, propane, i-butane, n-butane, i-pentane, n-pentane, etc.), that are present in NGL at various temperatures (**TABLE 2**). To calculate the solubility of water in NGL, the water saturation values of pure components are multiplied with the weight fraction of the components in NGL.

NGL water saturation value calculation. For a mixture of organic liquids like NGL, an average saturation value is calculated from the weight fractions and saturation values of the pure components by using Eq. 1:

$$\text{Average Cs} = \Sigma_{i=0}^{n} Xi(\text{Cs})i \quad (1)$$

where:

Xi = Weight fraction of the *ith* component

$(\text{Cs})i$ = The saturation concentration (ppmw) of the *ith* component

n = The total number of components.

TABLE 3 illustrates the steps followed to calculate average saturation values for each component of NGL sampled from the sample point GPP-1. **TABLE 4** summarizes the water saturation value of NGL sampled from various locations.

Water in NGL may exist under different forms, such as dissolved, emulsified or free water state. Once the water content increases above its solubility limit, then emulsified or free water will appear in the NGL. **FIG. 3** shows the water solubility curve of NGL sampled from the facility GPP-1, containing 1.28% of CH_4 plus 60.04% of C_2H_6 plus 24.45% of C_3H_8 plus 3.75% of i-C_4H_{10} plus 6.84% of n-C_4H_{10} plus 2.14% of C_5H_{12} plus 1.24% of C_5H_{12} plus 0.17% of n-C_6H_{14} plus 0.09% of n-C_7H_{16} with temperature ranges from 0°C to 50°C. In the region below the curve (Cs line), the soluble water alone can exist; above the line, both free water and soluble water can co-exist.

As shown in **FIG. 3,** once the water content of the sample increases above its solubility limit, the free water will appear along with NGL. **FIG. 3** shows that the solubility of water in NGL at 30°C is 215.6 ppmw. If the total water content rises above this value (at 30°C), then free water formation will occur.

TABLE 6. Total water contents at the outlet of separators

Facility	Sample temperature, °C	Solubility (soluble water), ppmw	Water content measured by KFT, ppmw	Free water content, ppmw
PS-1 Sample 1	47	405	3,000	2,594.5
PS-1 Sample 2	45	371.7	2,876	2,504.3
PS-1 Sample 3	40	288.4	339	50.6
PS-1 Sample 4	46	388.4	2,987	2,598.6
PS-1 Sample 5	48	421.7	1,012	590.3

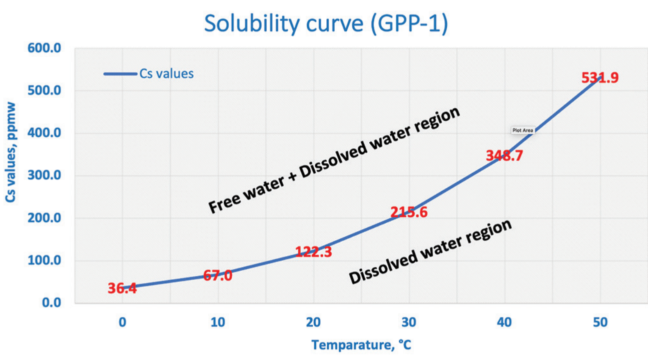

FIG. 3. Water solubility curve of GPP-1.

For example, if the sample contains a total water content of 250 ppm at 30°C, then 34.4 ppmw of water will be in the form of free water and 215.6 ppmw will be in the soluble form. As shown in **FIG. 3,** once the temperature rises, the solubility of the water NGL will also increase, and vice versa. The solubility limit of water in NGL depends on the temperature—i.e., the higher the temperature, the higher the solubility of water in NGL. **FIG. 4** represents the solubility curve of samples from other sample points.

Sample from downstream of inlet separators. The condensate from downstream of the inlet separators will contain a high water content, including free and soluble water. One sample point was selected for this study at different periods of time, and the results are summarized in **TABLE 6.** To calculate the Cs values, the composition was obtained from gas chromatography analysis. As shown in the table, total water concentrations were varied from a few hundred ppmw up to 0.3%. For illustration, the recorded temperature of Sample 1 is 47°C; at this temperature, the total water content measured by KFT is 3,000 ppmw. The solubility curve shows that the soluble water content of this sample will be 405.5 ppmw, and the free water content will be 2,594.5 ppm.

FIG. 5 shows the solubility curve of the sample from PS-1. In this sample point, it is recommended to design a facility to remove the free water by additional cooling upfront so that it can remove any free water early on.

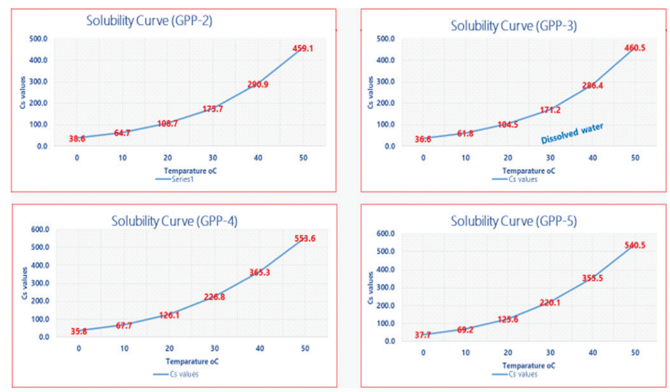

FIG. 4. Water solubility curves of samples from GPP-2, GPP-3, GPP-4 and GPP-5.

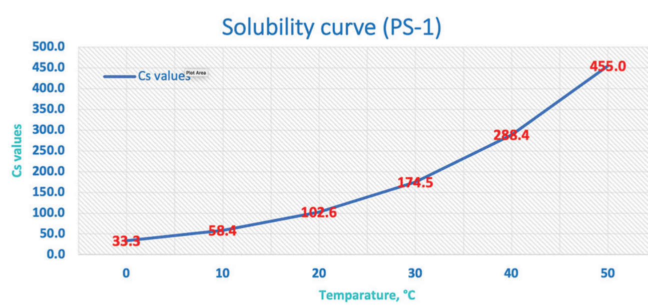

FIG. 5. Solubility of water in NGL vs. temperature

Recommendations. This study concluded that the solubility of water in NGL at varying process temperatures can be easily calculated, based on the NGL composition and the water saturation values of the pure components present.

The solubility curve can guide engineers in understanding the importance of improving the existing condensate dehydration step and the selection of the appropriate technology or technologies for processing NGL. From the solubility curve, it can be seen that if the plant engineer can maintain a total water concentration of 10 ppmw or below in processed NGL, then free water formation at the pipeline will not occur even if the ambient temperature of the pipeline goes to 0°C. **GP**

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LITERATURE CITED

Complete literature cited available online at www.GasProcessingNews.com.

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Optimize throughput of gas pipelines and facilities with intelligent software

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A drop of liquid. An ice formation. So small and yet so potentially costly and damaging to the operations of gas facilities and pipelines. Unwanted hydrates or liquids can clog pipes, reduce the process flow and damage equipment.

Gas processing technologies depend on understanding the phase envelope of the process fluid to operate efficiently and safely. Process engineers implement their designs to manage the worst anticipated phase envelope, normally resulting in less-than-optimal throughput and unnecessary energy use for typical or normal operating conditions.

To deal with these challenges, intricate calculations are needed and control actions must happen in real time.

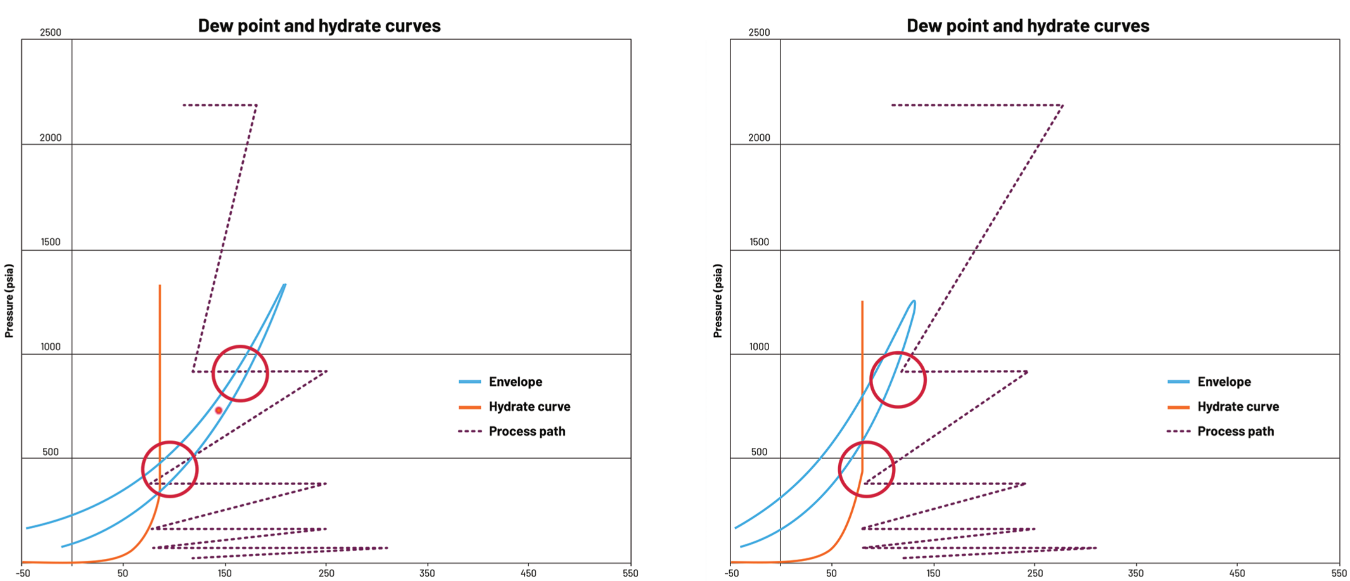
Most operators adjust controls to operate the plant or the pipeline a safe distance from the formation of hydrates or well into the phase the process needs to operate on. This is accomplished by injecting chemicals, increasing the temperature of the gas with heaters, reducing temperatures far below what is needed, or operating well below the design capacity. The first three methods increase the operational cost and indirectly reduce throughput. The last one directly reduces throughput. Consequently, opportunities to improve yield and efficiency dissipate.

All this in an industry era marked by falling budgets for capital and operational expenditures (CAPEX and OPEX). While this is true at many gas facilities and pipelines operating today, it no longer has to be that way.

Increasing throughput and reliability.

Until recently, no solution existed for the control system to adapt operations based on real-time conditions and compositions. Historically, process engineers calculated processing fluid simulations using spreadsheets on a desktop computer and shared them with the control engineers to implement. Those calculations helped ensure the plant operated in safe operating conditions, based on the expected range of the compositions of the process fluid. However, these calculations are based on safeguarding against worst-case scenarios, such as extreme temperatures or the poor quality of a raw material, and most of the time these facilities operate under much better circumstances.

Using intelligent throughput optimization software, that simulation capability can become a real-time control system activity to help prevent reliability issues and increase throughput. With insights provided by the software as the compo-



Facility failing to operate in the right phase | **Facility maintaining correct phase using Throughput Optimization**

FIG. 1. The phase envelope and hydrate curves for a refrigeration process. Both graphs show the same facility.

sition of the fluid being processed changes, the control strategy can be adapted to that change.

FIG. 1 shows the phase envelope and hydrate curves for a refrigeration process. Both graphs show the same facility. In the left graph, the facility is operating in an inappropriate state, with hydrates forming in one refrigeration cycle and the fluid becoming all liquid in another cycle. In the right graph, intelligent throughput optimization software is operating, providing closed-loop control to the control strategy to avoid hydrates and achieve the correct amount of cooling without causing liquids to form when they are not wanted.

With throughput optimization software, the shape and location of the hydrate curve and phase envelopes are known in real time. The process operation can be moved to the correct side of both physical entities. The adjustments can take place either through direct closed-loop control with the control system, or the software advises the operator through an human-machine interface (HMI) and allows the operator to adjust the control system.

The software helps the facility on the right increase reliability through the prevention of shutdowns, reducing energy by optimizing the refrigeration cycle and not exceeding what is needed. Additionally, by being able to operate efficiently close to the phase envelope, the facility's productivity increases.

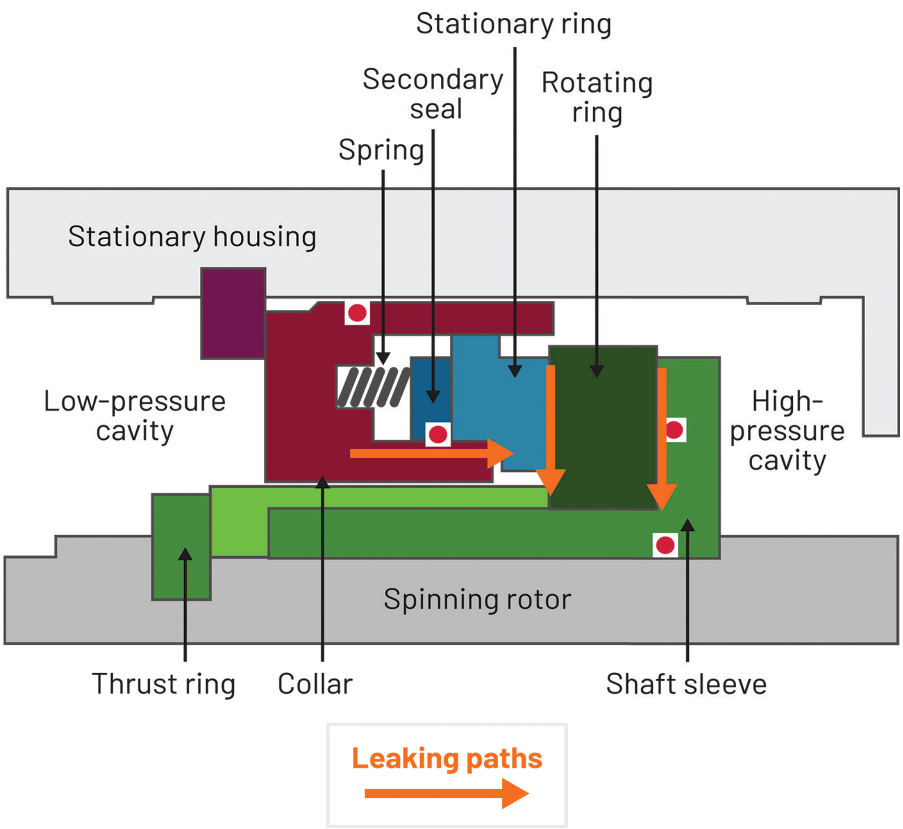


FIG. 2. A cutaway of a centrifugal compressor on a natural gas pipeline or processing plant that uses dry gas seals.

How throughput optimization works.

In throughput optimization, the targeted software application communicates with the controller running the plant or facility to provide a physics-based way to optimize the processing of fluids to avoid process upsets, and solves issues caused by failing to operate in the proper phase and/or the formation of hydrate or ice particles. As a result, the automated tuning control solution allows operations engineers to easily and reliably control the process to optimize operations based on real-time conditions and composition. This can measurably reduce energy costs in an array of applications, including:

- Avoiding hydrate formation in piping as the oil and gas comes out of the ground
- Ensuring liquids do not form in the piping as product flows to its consumption destination
- Monitoring the temperature in sulfur recovery units (SRUs).

For example, **FIG. 2** shows a cutaway of a centrifugal compressor on a natural gas pipeline or processing plant that uses dry gas seals. A microscopically thin film of gas provides lubrication and separation for the rotating elements. During startup and shutdown—particularly quick-stop and sudden shutdowns—the pressure and temperature of the gas film may move into the liquid phase envelope. A single, microscopic drop of fluid in the dry gas seal can destroy the seal.

Throughput optimization can identify if the potential for this exists, and, if necessary, operate a heater on the dry gas seal feed to avoid the issue.

For another use-case scenario, consider an SRU in which hydrogen sulfide (H_2S) is converted to elemental sulfur using the Claus process. The temperature of the furnace in the SRU is critical to proper and continuous operation but is difficult to measure due to the high temperatures, which consequently also rapidly destroy thermocouples. Throughput optimization can determine temperature based on composition and flowrate of inputs to the furnace.

The software can adjust automatically or alert an operator via a user-interface showing phase envelope and hydrate curve to make adjustments. **FIGS. 3–5** show how the software demonstrates the value of understanding the phase envelope and hydrate curve in operating a thermal natural gas separation system:

- **FIG. 3:** Without intelligent throughput optimization control, the process conditions are crossing the hydrate curve. Without action, hydrates will form, eventually causing vessels and piping to plug and the process to be shutdown.
- **FIG. 4:** Without intelligent throughput optimization control, the process conditions are so close to the vapor saturation curve that little if any liquid is forming, minimizing the effectiveness of the separator.
- **FIG. 5:** With closed-loop intelligent throughput process control, the

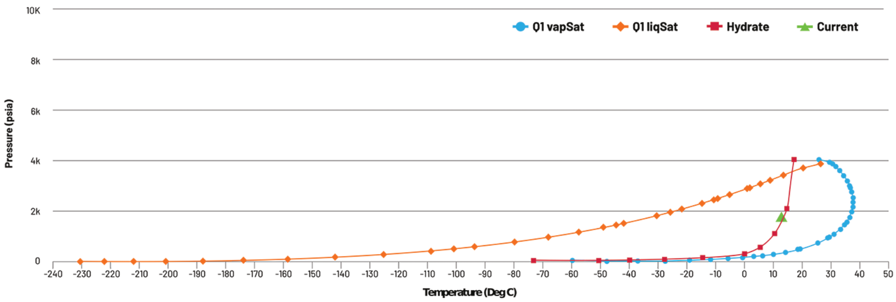


FIG. 3. Without intelligent throughput optimization control, the process conditions are crossing the hydrate curve.

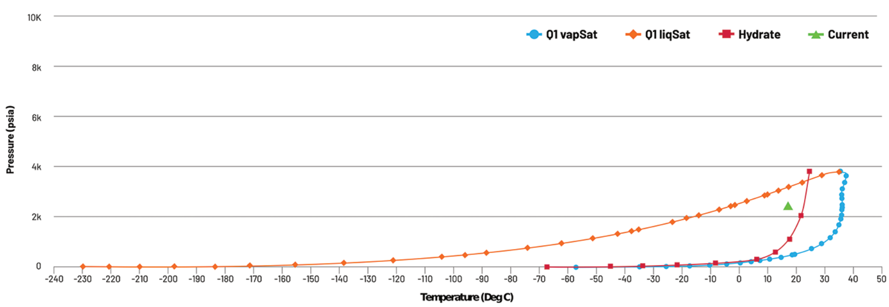


FIG. 4. The process conditions are so close to the vapor saturation curve that little if any liquid is forming, minimizing the effectiveness of the separator.

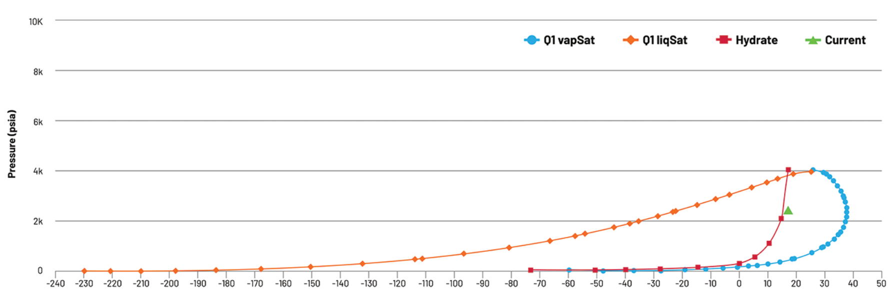


FIG. 5. With closed-loop intelligent throughput process control, the process is operating at an optimal point in the phase envelope for liquid dropout.

process is operating at an optimal point in the phase envelope for liquid dropout.

Real-world results. By gaining real-time insights from the two-phase envelope, producers can:

- **Increase throughput**—By understanding where the process is operating in relation to the phase envelope and hydrate curves, processes can be tuned to produce more output.
- **Increase reliability**—By understanding the phase envelope and hydrate curve, facilities can be operated to avoid mixed-phase operation and avoid hydrates, which leads to stable operation and less wear and tear on equipment.
- **Reduce energy use**—By understanding hydrate temperatures and phase envelopes, heater, refrigeration and compression use can be optimized, reducing energy use.
- **Reduce emissions**—Understanding process states results in a more stable facility, reducing emissions generated during instabilities and providing the inputs to optimization for minimum emissions.

Case studies. A processor shipping propane entrained in a natural gas pipeline in North America had conservatively based calculations on the lowest gas temperature of about 20°F (–7°C). This helped them gauge how much propane could be injected into the natural gas before liquid could form and collect in the low points of the pipelines, and thereby threaten compressor performance.

However, the temperatures rarely sank so low, which meant they were missing the opportunity to earn more yield. Since implementing the new simulation software, the company can use real-time, composition-based physical constants for closed-loop control and automatically set the propane injection limit factoring in the temperature. As a result, the producer can ship approximately 20% more propane throughput 90% of the year.

In another case, facility operators ran heaters around the clock to keep hydrates from forming in their natural gas piping. Throughput optimization simulation software, however, pinpointed the risks of hydration forming as next to none. This allowed the operators to reduce the use of heaters from 100% of the time to only 5%, saving more than $ 1 MM/yr in fuel costs.

Takeaway. As gas pipeline and facilities operations face mounting demands to increase throughput and reliability, producers are looking for pragmatic, cost-effective ways to improve production and efficiency. By using real-time automated tuning control solutions, producers can more effectively control processing fluids, while mitigating risks to processes, equipment and personnel. As a result, producers can maximize capacity use and better maintain operational safety and total costs. **GP**

ROSS OTTO focuses on pipeline control and optimization as a Global Engineering Manager and packaged solution champion at Sensia, a joint venture between Rockwell Automation and Schlumberger. Mr. Otto's leads an innovation team adapting existing technologies to specific industry challenges in optimization, maintenance and emissions reduction. He previously served as a technical specialist and program manager for Rockwell Automation. Before coming to the automation world, Mr. Otto spent many years designing and delivering pipeline SCADA, leak detection and optimization systems. With more than 35 yr of experience, he is a trusted source for pipeline automation and optimization in the oil and gas industry.

Customize selection of hydrocarbon dewpoint control technology

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The purpose of a hydrocarbon dewpoint control unit is to remove hydrocarbon liquids (NGL) from natural gas so that the hydrocarbon dewpoint specifications of the sales gas (local requirement is 0°C at any pressure) can be met. A challenge for plant designers is to choose the most suitable technology for hydrocarbon dewpoint control. The technology must be technically and economically capable of continuous, trouble-free operation, with optimum outcome. It must have the flexibility to adjust the water and hydrocarbon dewpoints of the sales gas to the required pipeline specification on maximum and turndown conditions without affecting the performance of the plant, even in the worst-case scenario.

The selection of appropriate technology for hydrocarbon dewpoint reduction must consider several factors. It is imperative for the plant designer to understand the key factors that contribute to the choice of the best technology for the plant, particularly at the conceptual design stage.

Three hydrocarbon dewpoint control technologies are commonly used: Joule-Thomson (JT) valve technology, turboexpander technology and mechanical refrigeration technology. Significant research is available on each type of technology independently, but a comparative analysis must be performed for each technology in relation to the others based on key impacting factors, as the optimum performance of the plant relies on the proper selection of technology.

This comparative process analysis will ascertain the application and pros and cons of each dewpoint control process and technology with the objective of selecting the most suitable process for the specified gas treatment plant. The ultimate selection is based on the financial viability of the process schemes and the technical comparison where the impact of the contributing factors is identified.

Process simulations are prepared for all three technologies with results to show the optimum scheme, along with other contributing factors in the selection of technology. The key contributing factors considered for comparison are complexity of operation, turndown, reliability and availability, pressure drop, sales gas specifications and others.

Mechanical refrigeration technology.

In the mechanical refrigeration scheme, gas from the upstream processing unit is routed to the gas/gas heat exchanger that provides a means of heat exchange between the raw gas and the low-temperature separator (LTS) gas; therefore, the precooling of the raw gas is achieved in this heat exchanger. As the temperature drops, some condensation occurs downstream of the gas/gas exchanger. The gas is then passed through the chiller to further reduce its temperature, which promotes additional condensation.

The combined stream [gas, condensed hydrocarbon and the ethylene glycol (EG)/methanol solution] is then introduced in the LTS to segregate the gas, liquid hydrocarbons and the EG/methanol solution. After the separation of NGL, gas from the LTS will have dewpoint, as per pipeline specifications. This dewpoint-specified gas is then routed through the gas/gas heat exchanger, as previously discussed, to exchange heat with the raw gas and exit as sales gas.

Cooling of the raw gas in the chiller is achieved by using propane as refrigerant for heat exchange. Propane gains energy and vaporizes in the chiller after heat exchange. Vaporized gas is routed to the refrigeration unit, where it is condensed after being recompressed in the refrigerant compressor and routed to the refrigerant accumulator. Propane fill/makeup in the refrigerant loop is carried out at the refrigerant accumulator. Compressed propane is finally cooled/chilled at the required temperature by pressure letdown across the valve. Chilled propane is then introduced back into the chiller for heat exchange.

NGL from the LTS is first passed through the multipass exchanger to raise its temperature above the hydrate formation temperature before being routed to NGL storage.

Glycol solution (EG) or methanol is injected in raw gas at the inlet of the gas/gas exchanger to avoid possible hydrate formation at low operating temperature. Rich EG/methanol solution collected from the LTS is routed to the glycol/methanol regeneration system, which, after processing, recycles back lean EG/methanol solution to the dewpoint reduction unit (**FIG. 1**).

Phase envelope. Proprietary simulation steady-state software[a] is utilized to simulate the hydrocarbon dewpoint control unit. **FIG. 2** shows the phase envelope for lean sales gas after processing through a mechanical refrigeration-based hydrocarbon dewpoint control unit.

Application pros and cons. The application of the mechanical refrigeration processing route has a number of benefits and challenges:

- Although this technology is conventional and can produce the required sales gas hydrocarbon dewpoint specification, it is not suitable at higher operating pressures, as more cooling is required at high operating pressures to extract the same amount of heavier component as compared to relatively low or moderate pressures

- If adequate pressures are not available at the feed of the hydrocarbon dewpoint control unit, then mechanical refrigeration can be used for the elimination of hydrocarbons from raw gas
- The availability and reliability of the unit is lower as compared to other units because of rotating equipment like the refrigerant compressor and the refrigerant cooler (air cooler); however, this can be mitigated with scheduled maintenance and proper monitoring
- Pressure drop in the gas path of the chiller and gas/gas exchanger is low; therefore, sales gas compression is not required
- A refrigerant is required for first fill and makeup due to losses from the system, which adds to OPEX; the costs of other components of the refrigeration unit add to CAPEX.

Turboexpander technology. A turboexpander scheme is used where a substantial amount of C_3+ components are available in the feed gas stream to be extracted. In the turboexpander scheme, gas from the upstream processing unit is routed to the gas/gas heat exchanger, which provides a means of heat exchange between the raw gas and the LTS gas; therefore, the precooling of the raw gas is achieved in this heat exchanger.

As the temperature drops, some condensation occurs downstream of the gas/gas exchanger, and the combined stream (gas, condensed hydrocarbon and the EG/methanol solution) is then introduced in the cold separator to segregate the gas, liquid hydrocarbons and the EG/methanol solution. Outlet gas from the cold separator is then isentropically expanded in the expander section of the turboexpander.

At the suction nozzle of the turboexpander, the ethylene glycol/methanol solution is injected as a fine mist to avoid possible hydrate formation in the expander and at the downstream system. The decrease in temperature due to expansion promotes further condensation. The combined stream (gas, condensed hydrocarbon and the ethylene glycol/methanol solution) is then introduced in the LTS to segregate gas, liquid hydrocarbons and the ethylene glycol solution. After separation of NGL, gas from the LTS will have dewpoint as per pipeline specifications; however, it still requires compression. This dewpoint-specified gas is then routed through the gas/gas heat exchanger to exchange heat with raw gas, followed by compression through the turboexpander-driven sales gas compressor up to the required sales gas pressure.

NGL from the LTS and cold separator is first passed through the multipass exchanger to raise its temperature above the hydrate formation temperature before being routed to NGL storage.

EG or methanol is injected in the raw gas at the inlet of the gas/gas exchanger to avoid possible hydrate formation at low operating temperature. Rich EG/methanol solution collected from the LTS and the cold separator is routed to the glycol/methanol regeneration system, which, after processing, recycles back lean ethylene glycol solution to the dewpoint reduction unit (FIG. 3).

Phase envelope. Proprietary simulation steady-state software[a] is utilized to simulate the hydrocarbon dewpoint control unit. FIG. 4 shows the phase envelope for lean sales gas after processing through the turboexpander-based hydrocarbon dewpoint control unit.

Application pros and cons. The application of the turboexpander technology processing route has a number of benefits and challenges:

- Turboexpander-based dewpoint reduction technology is mostly used in cryogenic process—i.e., where

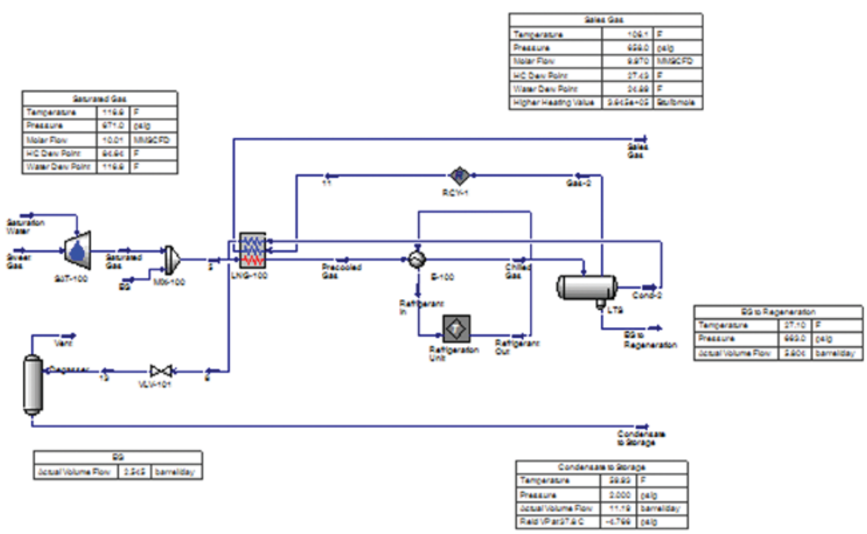


FIG. 1. Process scheme for mechanical refrigeration technology.

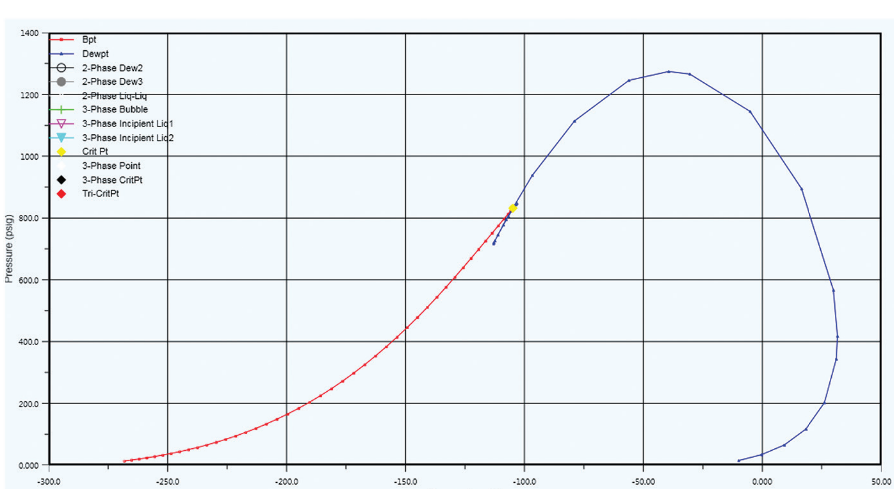


FIG. 2. Phase envelope for lean sales gas after processing through a mechanical refrigeration-based hydrocarbon dewpoint control unit.

LPG or high recoveries of C_2 and C_3+ components are required
- The major benefit of the turboexpander scheme is the utilization of expansion energy for the recompression of gas; therefore, the energy requirement for the external sales gas compressor is reduced
- The existence of complex rotating equipment (expander and compressor) demands maintenance and operation expertise, which impacts the availability and reliability of the plant
- Being complex and high in capital cost, turboexpander technology is used for high and constant feed flowrates, and it is difficult to control and turndown
- When used for cryogenic process, almost fully dehydrated gas is required by turboexpander technology to prevent hydrate formation, as LTS temperature will drop in cryogenic range; therefore, a solid-bed system or a silica gel-based dehydration unit is required instead of a glycol dehydration unit to dehydrate the gas completely before processing it through a turboexpander
- With the use of this technology, very low hydrocarbon dewpoint can be achieved.

JT valve technology. In principle, JT valve technology utilizes the Joule-Thomson effect to achieve the required cooling for separation of heavier hydrocarbons, thereby attaining dewpoint reduction in gas at the outlet of the unit. Gaseous hydrocarbons are cooled by throttling its flow through the JT valve, initiating rapid expansion. This phenomenon is known as the Joules Thomson effect. The quantity of NGLs recovered is a function of the gas composition, pressure, temperature and pressure drop across the JT valve.

Gas from upstream of the processing unit is routed to a gas/gas heat exchanger, which provides a means of heat exchange between the raw gas and LTS gas; therefore, precooling of the raw gas is achieved in the heat exchanger. As the temperature drops, some condensation occurs downstream of the gas/ gas exchanger.

Outlet gas from the cold separator is then adiabatically expanded through a pressure control valve (JT valve).

The decrease in temperature due to expansion promotes further condensation. The combined stream (gas, condensed hydrocarbon and the EG/methanol solution) is then introduced in the LTS to segregate gas, liquid hydrocarbons and the ethylene glycol solution.

After the separation of NGL, gas from the LTS will have dewpoint as per pipeline specifications; however, it still requires compression. This dewpoint-specified gas is then routed through the gas/gas heat exchanger to exchange heat with raw gas, followed by compression through the sales gas compressor up to the required sales gas pressure.

EG or methanol is injected in raw gas at the inlet of the gas/gas exchanger to avoid possible hydrate formation at low operating temperature. Rich EG/methanol solution collected from the LTS is routed to the glycol/methanol regeneration system, which, after processing, recycles back lean ethylene glycol solution to the dewpoint reduction unit.

Condensate (NGL) from the LTS is first passed through the multipass exchanger to raise its temperature above the hydrate formation temperature before being routed to NGL storage (**FIG. 5**).

Phase envelope. Proprietary simulation steady-state software[a] is utilized to simulate the hydrocarbon dewpoint control unit. **FIG. 6** reveals the phase envelope for lean sales gas after processing through a JT valve-based hydrocarbon dewpoint control unit.

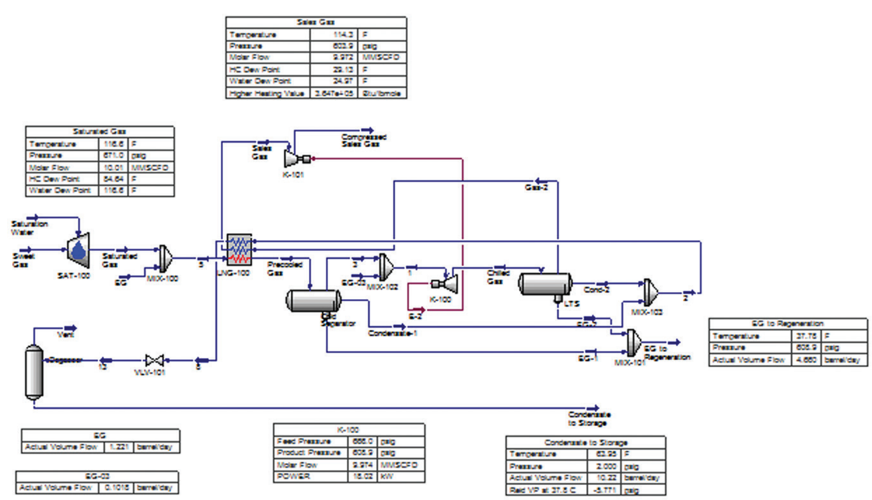


FIG. 3. Process scheme for turboexpander technology.

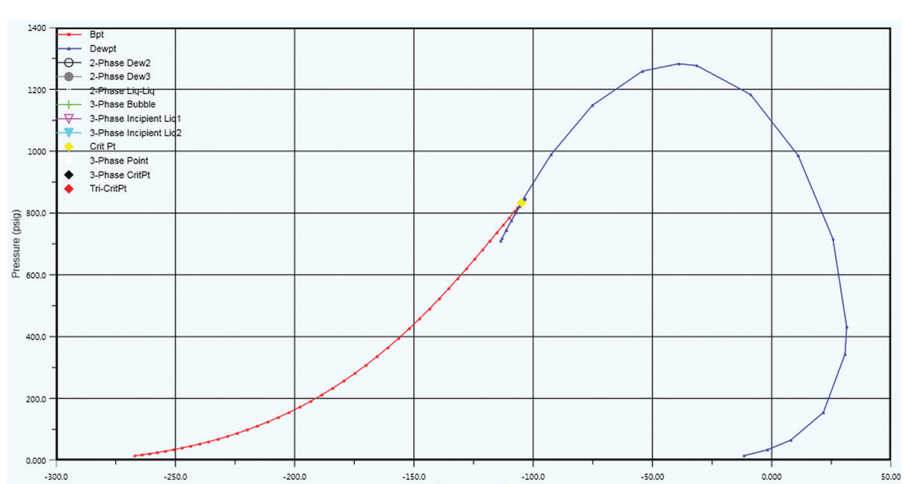


FIG. 4. Phase envelope for lean sales gas after processing through the turboexpander-based hydrocarbon dewpoint control unit.

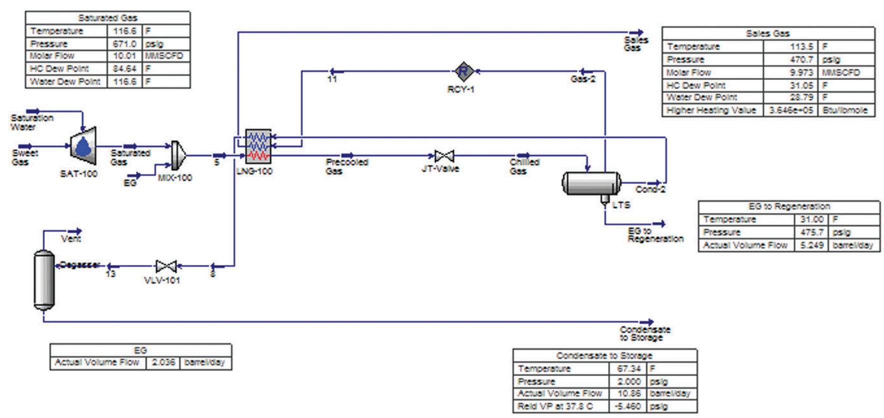


FIG. 5. Process scheme for JT valve technology.

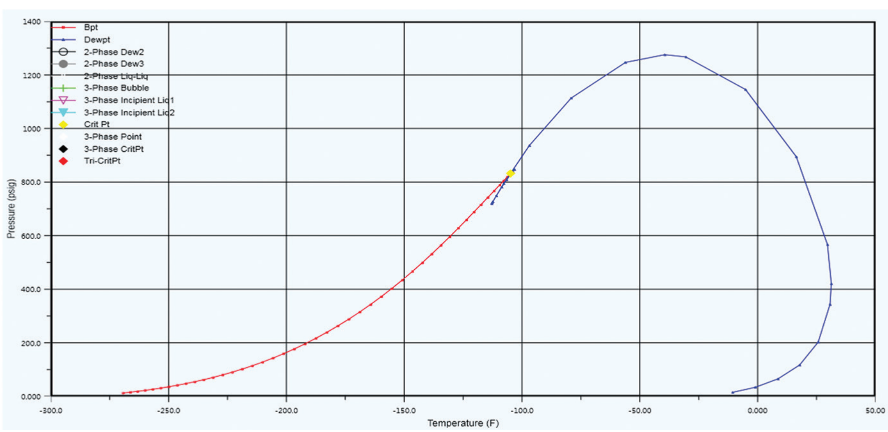


FIG. 6. Phase envelope for lean sales gas after processing through a JT valve-based hydrocarbon dewpoint control unit

Application pros and cons. The application of the JT valve technology processing route has a number of benefits and challenges:

- For small, high-pressure feed gas streams, JT valve is the technology of choice to yield on-spec hydrocarbon dewpoint sales gas
- High turndown ratio can be achieved with this technology
- It is a self-refrigeration process; no external cooling medium is required
- It requires high pressure drop, and pressure that is dropped across the JT valve cannot be recovered; energy is required to recompress the on-spec dewpoint gas as per sales gas injection pressure
- JT valve technology has the capability to operate on high turndown, with ease of operation.

Technical comparison of technologies.

Key technology features are compared in the following sections.

Operation. In the absence of complex rotating equipment like a turboexpander or a refrigeration compressor required for cooling, JT valve technology has a simple and flexible operation.

Turndown ratio. JT valve technology has the highest turndown ration among the proven technologies.

Turboexpander technology has the lowest turndown ratio among all the options, as a centrifugal compressor is an integral part of the turboexpander.

With the use of a screw compressor as part of the mechanical refrigeration unit, turndown almost equal to that of JT valve technology can be achieved.

Reliability and availability. JT valve technology has the highest reliability among all considered options, whereas turboexpander technology has the lowest reliability due to the number of rotating equipment.

Pressure drop. JT valve technology requires high pressure drop for cooling and requires the highest quantity of fuel gas for sales gas compression.

Mechanical refrigeration technology requires the lowest pressure drop and lowest quantity of fuel gas; however, this saving in energy is partially offset by running the refrigeration compressor.

Turboexpander technology requires a higher pressure drop for expansion than a mechanical refrigeration unit, but the expander and compressor partially compensate for this.

Recompression requirements. Recompression is required for JT valve technology, whereas other technologies do not require recompression, since their pressure letdown is low.

Feed gas requirements. Turboexpander technology can be used for high feed gas flowrates; however, other technologies can be used for a wide range of feed gas flowrates.

Sales gas specification. Sales gas specifications can be met through all of the compared hydrocarbon dewpoint technologies.

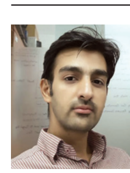
Recovery. Sales gas recovery with JT valve technology and NGL recovery with mechanical refrigeration technology are highest; however, simulation operating parameters used will have an effect on these recoveries.

Takeaway.

The basis application for each type of hydrocarbon dewpoint technology should be considered against the specific requirement of the gas to be processed. The capital cost of each technology, which is not considered in this article, should also be taken into account along with the limitations and suitability of the technologies discussed in this article. The better the technology selection, the more optimum will be the results. **GP**

NOTES

[a] HYSYS



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Japan to develop standard emissions measure for LNG

Japan Oil, Gas and Metals National Corp. (JOGMEC) intends to develop a global standard for measuring greenhouse gas emissions in the LNG value chain. The LNG sector is facing increasing pressure to cut emissions of greenhouse gas, including methane, to help tackle climate change, but calculation methods vary by country and company.

The company aims to verify the methodology using actual data from LNG plants in the near future. The methodology will be internationally comparable to promote emissions cuts in every phase of LNG production and distribution, and is aimed at making new developments cleaner so that new projects can secure financing.

Shell and Enbridge to produce RNG, H_2

Canadian pipeline operator Enbridge Inc. has signed partnerships with Royal Dutch Shell and Vanguard Renewables to make low-carbon fuels, seeking to tap into sales to companies that aim to lower their greenhouse gas emissions.

Enbridge will buy 2 Bft^3y of renewable natural gas (RNG) from Vanguard and collaborate with Shell on potential green and blue hydrogen production. Companies that buy RNG from Enbridge would collect the offsets associated with decarbonization. Enbridge, which set emissions-reduction targets in late 2020, hopes to be a net-zero emitter by 2050.

Alaska LNG project to help Asia cut CO_2

Alaska Gasline Development Corp. (AGDC) said that its proposed LNG export project would reduce greenhouse gas emissions in Asia by allowing power generators to use a cleaner fuel than coal.

The state-owned LNG and pipeline developer released a study concluding that overall greenhouse gas emissions from Alaska LNG natural gas would be 50% less than burning Chinese regional coal, reducing CO_2 emissions by 77 metric MMtpy. The study also showed that Alaska LNG had a lower greenhouse gas intensity than other LNG export projects on the U.S. Gulf Coast and Australia.

AGDC is developing the $38.7-B Alaska LNG project, which includes a liquefaction facility on the Kenai Peninsula in southern Alaska and a proposed, 807-m (1,299-km) pipeline that would move gas stranded in northern Alaska across the state.

AGDC is not looking to build Alaska LNG itself, but instead work with parties that can build a gas treatment plant in northern Alaska, the pipeline and liquefaction facility. AGDC has already lined up parties to lead the pipeline and gas treatment plant, and is still looking for a party to lead construction of the liquefaction plant.

To help the project move forward, the U.S. Congress has included loan guarantees worth $26 B for the Alaska LNG project in the latest infrastructure bill.

Shell and BASF to collaborate on CCS technology

Shell and BASF are collaborating to accelerate the transition to a world of net-zero emissions. To this end, both companies worked together to evaluate, de-risk, and deploy BASF's Sorbead® Adsorption Technology for pre- and post-combustion carbon capture and storage (CCS) applications. The Sorbead Adsorption Technology is used to dehydrate CO_2 gas after it has been captured by Shell's carbon capture technologies, such as ADIP Ultra or CANSOLV.

The adsorption technology has several advantages for CCS applications: Sorbead, an aluminosilicate gel material, is acid resistant, has high capacity for water and regenerates at a lower temperature vs. activated alumina or molecular sieves. Furthermore, Sorbead Adsorption Technology ensures the treated gas is free of glycol and will meet stringent pipeline and underground storage specifications. Customers also benefit from long life, operational turndown flexibility and immediate on-spec gas at startup.

The Sorbead Adsorption Technology is now in Shell's portfolio for use in the numerous CCS projects around the world to achieve their Powering Progress strategy.

Technip Energies, Shell test the latest Cansolv CO_2 capture technology improvements

Technip Energies and Shell Catalysts & Technologies unveiled that their jointly developed improvements on the Cansolv CO_2 capture technology are being tested in a pilot plant campaign at Fortum Oslo Varme's waste-to-energy plant.

In response to the increased global interest in CCS, the technologists and engineers of the two companies are working closely to bring continuous improvements to the Cansolv CO_2 capture system's process design, efficiency and costs. These efforts are to ensure improved affordability and aid wide scale deployment of carbon capture solutions by their clients.

The test campaign will entail different test phases that will support the extension of the related improvements. This is the second test campaign conducted by this collaboration. In the first campaign, the low volatility and amine emissions of the DC-103 solvent used in the process was demonstrated, as well as its low absorption energy and solvent degradation.

FuelCell Energy and ExxonMobil extend agreement for carbon capture technology

FuelCell Energy, Inc., has signed a 6-mos extension with ExxonMobil to continue collaboration on carbonate fuel cell technology for the purpose of capturing CO_2 from industrial facilities and power generation.

The agreement will continue until April 30, 2022. The parties are discussing an ExxonMobil pilot in Rotterdam, the Netherlands, as well as potentially additional ExxonMobil or third-party locations, to deploy FuelCell Energy's carbonate fuel cell platform to capture CO_2 emissions. A decision on the Rotterdam project is expected in 2022, dependent on achieving technical milestones over the next 6 mos. In addition to pilot project deployments, FuelCell Energy and ExxonMobil are discussing the next phase of carbon capture development.

FuelCell Energy's technology will help capture CO_2 emissions from power generation and industrial exhaust streams. Together with ExxonMobil, the companies can scale and commercialize FuelCell Energy's carbon capture solution, one that captures carbon dioxide from various exhaust streams, while generating additional power, unlike traditional carbon capture technologies, which consume significant power.

FuelCell Energy's proprietary technology uses carbonate fuel cells to efficiently capture and concentrate CO_2 streams from industrial sources. Combustion exhaust is directed to the fuel cell, which produces power, while capturing and concentrating CO_2 for permanent storage. The modular design enables the technology to be deployed at a wide range of locations, which could lead to a more cost-efficient path for deployment of CCS.

Technology to reduce methane emissions in Barnett

TotalEnergies plans to deploy an innovative technology, developed by Qnergy, to significantly reduce methane emissions related to its operations in the Barnett gas field in the U.S. The solution uses a technology that converts methane-powered instrumentation to compressed air-powered instrumentation, thereby eliminating the release of methane to the atmosphere during the process.

During a successful pilot project at the Barnett site in March 2021, Qnergy's technology proved to be reliable, simple to install and easy to operate, allowing the elimination of up to 98% of the methane venting emissions related to instruments using natural gas.

Following successful additional tests, TotalEnergies has decided to install the new technology by deploying 100 units on the Barnett field in 2021 and 2022. The deployment of 300 additional units throughout the field will reduce methane venting emissions from pneumatic devices by approximately 7,000 tpy by the end of 2024.

JAX LNG completes first renewable LNG bunkering facility in the U.S.

JAX LNG, a small-scale LNG facility located along the St. John's River in Jacksonville, Florida, completed the first fueling of a marine vessel in the U.S. with a blend of LNG and renewable LNG (RLNG).

JAX LNG loaded the RLNG/LNG blend into the *Clean Jacksonville* bunker barge to fuel the *Isla Bella.* The *Isla Bella* is the world's first LNG-powered container ship and was put into service by TOTE Maritime Puerto Rico in 2015. Element Markets supplied the renewable natural gas (RNG) used to produce the RLNG via renewable thermal certificates (RTCs). Using RLNG to fuel marine vessels is a readily available pathway to net-zero emissions by 2050. RLNG's emissions profile as a maritime fuel is superior even to that of LNG, which already reduces greenhouse gas emissions by more than 25% over ultra-low sulfur diesel.

Decarbonization of the transport sector has greatly accelerated using regulatory incentives such as the alternative fuel tax credit, which encourages companies to adjust operations and make investments in assets that reduce carbon intensity.

Produced from the decomposition of organic waste, RNG is compatible with existing natural gas infrastructure, providing a practical and replicable source of energy that mitigates and repurposes carbon emissions. For this bunkering event, RTCs were matched to the physical LNG loaded into the *Clean Jacksonville* to create the RLNG/LNG blended product.

Axens, Arol Energy sign license agreement for Connect'In technology

Axens and Arol Energy have signed a license agreement for Axens' Connect'In® technology. This service offer will now be proposed by Arol Energy to its biomethane producer customers. By collecting and continuously processing industrial data, the digital tool provides performance monitoring, which can be used to provide optimization recommendations. The overview of the operation of the units allows customers to clearly identify improvement opportunities. They are then able to maximize the economic performance of the connected units by making regular adjustments to the operating parameters. By offering Connect'In® to its biomethane producer customers, Arol Energy enables them to maximize and secure the production of biomethane with the required qualities for its reinjection into the network.